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TECHNOLOGY IN ACTION

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PARTIAL GASIFICATION
of
NOVA SCOTIA COALS

An evaluation of; the technologies available to carry out partial gasification of Nova Scotia coals, the potential scenarios for the utilization of the technology and the economics of the application of the technologies.

Prepared by

Neil Stover, P.Eng.

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NOVA SCOTIA RESEARCH FOUNDATION CORPORATION
P.O. Box 790, Dartmouth, Nova Scotia, Canada, B2Y 3Z7
Tel: (902) 424-8670
Telex: 019-22719
Telecopier: (902) 465-7384

ABSTRACT

Processes that are suitable to partially gasify Nova Scotia coals in order to generate electricity or produce co-products (char, tars and gas) have been identified and evaluated.

Nova Scotia has active coal mines in the Cape Breton and Pictou County areas, all of which produce coals of the high volatile A bituminous rank.

Ash and sulphur concentrations vary from mine to mine and within the seam of each mine. The Cape Breton coals typically have sulphur in the 3% range and the Pictou coals in the <1% range.

A review of inactive and active processes that are capable of carrying out partial gasification of Nova Scotia coals has been carried out. The information regarding the inactive processes has been included to show perspective of the development of the technology. Numerous reactor designs have been developed to accommodate the caking properties of coals similar to those found in Nova Scotia.

The Vereinigte Elektrizitätswerke Westfalen (VEW) process in Dortmund, Germany, has been chosen for detailed technical and economic evaluation for generation of electricity. The VEW reactor is an entrained down flow design that processes 10 tonnes/h of a variety of coal types at 2 MPa and 1600°C. The economic analysis of a conceptual plant indicates that an 8% return on investment would require the levelized cost of electricity to be approximately 60 mils/kWh.

In the co-product category, the Institute of Gas Technology, Chicago, USA, process has been selected for more detailed technical and economic evaluation. The co-products are gas, that will be used to provide process heat, light and middle distillate oils that may be used as chemical feedstock, pitch that is targeted to be used as electrode binder and char which could be used in three applications: smokeless fuel, activated char for absorption and form coke to be used in metallurgical processes. The process gasifier is a fluidized bed that is topped with a circulating bed which operates at a solids handling rate of 45 kg/h. A 450 kg/h unit has been designed and construction will begin later this year. Conceptual processes would have major products of a char/coal blend utility fuel, form coke and smokeless fuel. The char/coal utility fuel option cannot be shown to be economically viable. The base case analysis of the smokeless fuel and form coke scenarios show return on investments of 5 and 13% respectively.

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DISCLAIMER

This report does not necessarily reflect the views of CANMET/Energy, Mines and Resources; the Cape Breton Development Corporation; the Nova Scotia Power Corporation and the Nova Scotia Department of Natural Resources, who co-sponsored this project, or their representatives. Neither the above parties nor their representatives make any warranty, express or implied, in respect of this report or its contents.

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ABBREVIATIONS

BC	British Coal
BP	Boiling point
BSR	Bench scale research
BSU	Bench scale unit
BTX	Benzene, toluene, xylene
C	Canadian
CBDC	Cape Breton Development Corporation
CB	Cape Breton, Nova Scotia
CEA	Canadian Electrical Association
CEGB	Central Electrical Generation Board
CFBC	Circulating fluidized bed combustor
COE	Cost of electricity
CR	Capital required
CRE	Cheltenham Research Establishment
CSIRO	Commonwealth Scientific Industrial Research Organization
CSR	Coke strength after reduction
DCF	Discounted cash flow
dif	Difference
EP	Average annual electricity production
EPRI	Electric Power Research Institute
ESP	Electro-static precipitator
f_a	ratio of aromatic to aliphatic carbon
FB	Fluidized bed
FC	Average annual fuel cost
FT	Fluid temperature, ash fusion
G	Giga (1E9)
GC	Gas chromatography
GTCP	Grimethorpe Topping Cycle Project
Hg	Mercury
HP	Horse power
HRSG	Heat recovery steam generation
HT	Hemispherical temperature, ash fusion
ID	Internal diameter
IGCC	Integrated gasification combined cycle
IGT	Institute of Gas Technology
IT	Initial temperature, ash fusion
KRW	Kellogg-Rust-Westinghouse
LCV	Low calorific value
LHV	Lower heating value
LLNL	Lawrence Livermore National Laboratory
LPG	Liquid petroleum product
LR	Lurgi-Ruhrgas
M	Million (1E6)
maf	Moisture and ash free
N	Normal (101 kPa & 0°C)
NDT	Net steam turbine output
NGT	Net gas turbine output

ABBREVIATIONS (Continued)

NSPC	Nova Scotia Power Corporation
OCC	Occidental Petroleum
OD	Outside diameter
OMC	Operating and maintenance costs
PBFBC	Pressurized bubbling fluidized bed combustor
PDU	Process development unit
PFB	Pressurized fluidized bed
PFBC	Pressurized fluidized bed combustor
pH	Negative of the log of the hydorgen ion concentration
PRU	Process research unit
PSFB	Pressurized spouted fluidized bed
ROI	Return on investment
ST	Spherical temperature, ash fusion
TGA	Thermogravimetric analyzer
USDOE	United States Department of Energy
VEW	Vereinigte Elektrizitätswerke Westfalen
VM	Volatile mater
WC	Working capital
wt	Weight
WV	West Virginia
yr	Year

1 INTRODUCTION AND BACKGROUND

The overall objective of this work has been to provide information regarding the application of partial gasification technology to Nova Scotia coals.

A technical and economic evaluation of coal partial gasification processes has been carried out. Gasification of coal occurs when it is heated to a temperature such that the chemical structure of the coal breaks down to form smaller molecules that are gases at ambient conditions. The gasification atmosphere must be reduced in oxygen concentration from that needed for complete combustion. When gasification is purposely carried out such that it leaves a portion of the coal as a solid char, it is termed partial gasification.

Environmental concerns and the high cost of coal mined in Nova Scotia result in the need to determine highly efficient alternatives for the conversion and utilization of the indigenous coal reserves.

The technology of partial gasification of coal has been primarily developed to achieve two ends:

- the production of electrical energy, and
- the production of value added products (co-products) in the form of gases, liquids and solids.

The application of partial gasification to electrical generation is carried out because of the promise of increased cycle efficiencies in comparison to conventional technology such as pulverized coal fired plants. Cycle efficiency may be defined as the fraction of the chemical energy in the coal that is converted to electrical energy available to the grid. Increased cycle efficiencies (based on the same fuel source) offer the potential of reduced emissions to the environment per unit of energy produced.

The motivation for the production of co-products is to increase the effective value of the coal by transforming it into products that are demanded in the market place such that they have greater value. The conversion of the coal to value added products improves the probability of economically mining and utilizing the coal.

The report begins with an Executive Summary which contains a reduced version of the more in-depth analysis that follows it.

2 EXECUTIVE SUMMARY

A technical and economic evaluation of coal partial gasification processes has been carried out. Gasification of coal occurs when it is heated to a temperature such that the chemical structure of the coal breaks down to form smaller molecules that are gases at ambient conditions. The gasification atmosphere must be reduced in oxygen concentration from that needed for complete combustion. When gasification is purposely carried out such that it leaves a portion of the coal as a solid char, it is termed partial gasification.

2.1 Objectives and Introduction

The objective of this work is to provide a body of information to coal producers and users regarding the status of the technology available to partially gasify Nova Scotia coals.

The technology of partial gasification of coal has been primarily developed to achieve two ends:

- the production of electrical energy, and
- the production of value added products (co-products) in the form of gases, liquids and solids.

The application of partial gasification to electrical generation is carried out because of the promise of increased cycle efficiencies in comparison to conventional technology such as pulverized coal fired plants. Cycle efficiency may be defined as the fraction of the chemical energy in the coal that is converted to electrical energy available to the grid. Increased cycle efficiencies (based on the same fuel source) offer the potential of reduced emissions to the environment per unit of energy produced.

The motivation for the production of co-products is to increase the effective value of the coal by transforming it into products that are demanded in the market place such that they have greater value. The conversion of the coal to value added products improves the probability of economically mining and utilizing the coal.

2.2 Highlights of Findings

The remainder of the executive summary contains a simplified and reduced version of the main body of the report.

2.2.1 Coal in Nova Scotia

Nova Scotia has two areas where significant coal recovery operations are active, Cape Breton and Pictou County. Table 2.1 lists some of the data for the active mines.

Table 2.1

Status of the Active Coal Mines in Nova Scotia

Mine/Seam	In Situ Reserves (10^6 tonnes)	Recoverable Reserves (10^6 tonnes)	Production (tonnes/day)
Prince/Hub	102	53	5,000
Lingan/Harbour	47	37	4,000
Phalen/Phalen	99	74	12,500
Westray/Foord	47	18	2,740

A new access tunnel is planned for the Prince mine. A conversion to retreat followed by closure of the mine is planned for Lingan. The Westray mine has started preliminary production.

All of the coals are classified as high volatile A bituminous (HvAb). Table 2.2 provides some typical values of the pertinent analysis of the coals on an in situ basis. Significant variations in some of the values shown in Table 2.2 can occur as the location of mining in the seam or the degree of cleaning is changed.

Table 2.2

Selected Analysis of Nova Scotia Coals

	Mine/Seam			
	Prince/Hub	Lingan/Harbour	Phalen/Phalen	Westray/Foord
Proximate Analysis (dry basis)				
% Ash	7	5	8	22
% Volatile Matter	37	37	38	28
% Fixed Carbon	56	58	57	54
% Moisture	6	1.5	2.5	8
Free Swelling Index	4	7	8	5
% Sulphur	4	2	2.5	0.8
Calorific Value, dry (MJ/kg)	31.4	33.9	32.6	30.6

The free swelling index (FSI) indicates the degree of swelling that the coal will undergo upon heating. This parameter is also indicative of a relative potential of problems caused due to the stickiness of the heated coal.

2.2.2 Review of Non-Active Processes

A review of inactive partial gasification processes, that were capable of processing Nova Scotia coals, was carried out to improve the perspective analysis of the active processes presented later.

Electrical Generation Process

The only process reviewed that was applied to electrical generation was the Lurgi-Ruhrgas (LR) in Germany.

Extensive testing of the LR system connected directly to a boiler furnace was done between 1957 and 1960. The final configuration of the LR III plant at Dorsten, Germany processed 10 t/h of dried coal operating typically at 920°C. The very fine grained portion of the char collected from various steps was blown directly into the boiler firing chamber through a refractory insulated line.

Co-Product Producing Processes

A number of reactor designs, including fluidized beds and entrained beds, were used in order to accommodate the caking (sticking) properties of the coals.

The condensable hydrocarbons, called tars, are comprised of products as described in Table 2.3.

Table 2.3

Description of the Condensable Hydrocarbon Fractions

Product	Description
Light Oils	The light oils are distilled from the tar and scrubbed from the gas stream and have a boiling point < 182°C.
Middle Oils	The middle oils are distilled from the tar in a temperature range between 182 and 310°C.
Heavy Oils	The heavy oils tar fraction has a distillation range between 310 and 400°C.
Medium-Soft Pitch	The medium-soft pitch is the untreated distillation residue from the tar fractionation.

The tars are primarily comprised of aromatic hydrocarbons (various complexes of benzene molecules). The concentration of heteroatoms (sulphur, oxygen and nitrogen) will be dependent on the characteristics of the coal, the gasifier design and its operational conditions. Heteroatoms typically reduce the stability of the tar by increasing the rate of re-polymerization which results in gum formation. Hydrotreating, the substitution of hydrogen for the heteroatoms, under conditions of elevated temperature and pressure must be carried out to reduce the heteroatom concentrations to acceptable levels. Increased heteroatom concentration therefore reduces the value of the tars.

Table 2.4 presents a summary of the inactive partial gasification processes that have been reviewed in the body of the report in more detail. Co-product yield data is given for the processing of HvAb coals that have similar characteristics to Nova Scotia coals. The process names are abbreviated as follows:

- Commonwealth Scientific and Industrial Research Organization (CSIRO)

- The Occidental Research Corporation Process (OCC)

- Lurgi-Ruhrgas (LR)

Table 2.4

Inactive Co-Product Process Reactor Design and Yields

Process Name	Reactor Type	Coal Process Rate (kg/h)	Yields (weight %, moisture, ash free (maf) basis)		
			Gases	Tars	Char
CSIRO	Fluidized bed	20	4	30	66
OCC	Down-flow entrained	91	6	30	64
LR	Twin-screw	500	13	21	66

2.2.3 Review of Active Processes

2.2.3.1 Electrical Generation

The partial gasification of coal to generate electricity would utilize the following general steps:

1. *Gasifier*: Coal is partially gasified resulting in the production of solids (char) containing mostly carbon and any inert material such as ash, and combustible gases such as carbon monoxide, hydrogen and methane. Some energy recovery may take place in this step which sees the generation of steam that is sent to the steam turbine.

2. *Combustor*: The char is sent to a combustor, typically a circulating fluidized bed at atmospheric or elevated pressure, where the remainder of the original chemical energy that was available in the coal is released during the combustion process. Steam is raised for use in a steam turbine.

3. *Gas Turbine*: The gases from the gasifier are combusted and expanded through a gas turbine. The hot turbine exhaust gases (mainly nitrogen, carbon dioxide and oxygen) are sent to the combustor.

4. *Steam Turbine*: Steam generated mainly in the gasifier heat recovery system and the combustor is sent to the steam turbine. This main steam flow along with other smaller amounts of steam generated in various heat recovery steps throughout the plant are used to generate electricity.

One of the advantages of the partial gasification process over the total gasification process is that the gasifier reactor costs are lower. The initial steps of volatilization (release of gases) and gasification (formation of gases) of coal are achieved relatively easily. As the gasification process progresses, the remaining char is more difficult to breakdown into gases. The conversion of the remaining carbon can be carried out much easier under the oxidizing combustion conditions seen in the char combustor. An optimum can be achieved between the gasifier and combustor sizing.

The optimization of the operation and sizing of the steam and gas turbines can also be carried out in conjunction with the combustor and gasifier sizing.

The optimum sizing of the gasifier, combustor and turbines results in the prediction of cycle efficiencies in the 45% range based on the coal HHV.

Table 2.5 summarizes the status of the processes of the major developers of partial gasification for electrical power generation.

Table 2.5

Companies Involved in the Evaluation of Partial Gasification of Coal As Applied to Electricity Generation

Process Developer	Reactor Type	Scale (kg/h)
British Coal	Spouted Bed	500
M. W. Kellogg	Entrained	42
Vereinigte Elektrizitätswerke Westfalen	Entrained	10,000

British Coal is actively developing what they call the "British Coal Topping Cycle". The spouted bed gasifier utilizes a fluidized bed variant where a spout of high velocity gas is introduced into the center of the bottom of a bed of solid material. The turbulent mixing caused by the high velocity spout prevents the sticking of coal particles which allows the processing of caking coals. British Coal has development programs in the area of pressurized spouted bed gasification, pressurized fluidized bed combustion (applicable to char combustion and steam generation), hot gas particulate removal (to protect gas turbine blades) and gas turbine materials (that are resistant to high temperature corrosion and erosion).

British Coal is proposing topping cycles utilizing either a pressurized bubbling fluidized bed char combustor (PFBC) (see Figure 2.1) or an atmospheric pressure circulating fluidized bed combustor (CFBC) (see Figure 2.2). A bubbling fluidized bed has a region of particles (approximately 60% by volume particles) that has gas moving upward through it at approximately 1.5 m/s with increasing diameter bubbles. The bed of material behaves as a fluid. A circulating fluidized bed has the gas moving through a region of decreasing solid concentration (20% to less than 5% particles) at velocities in the range of 5 m/s.

British Coal is projecting the operation of a demonstration scale plant by approximately 1995.

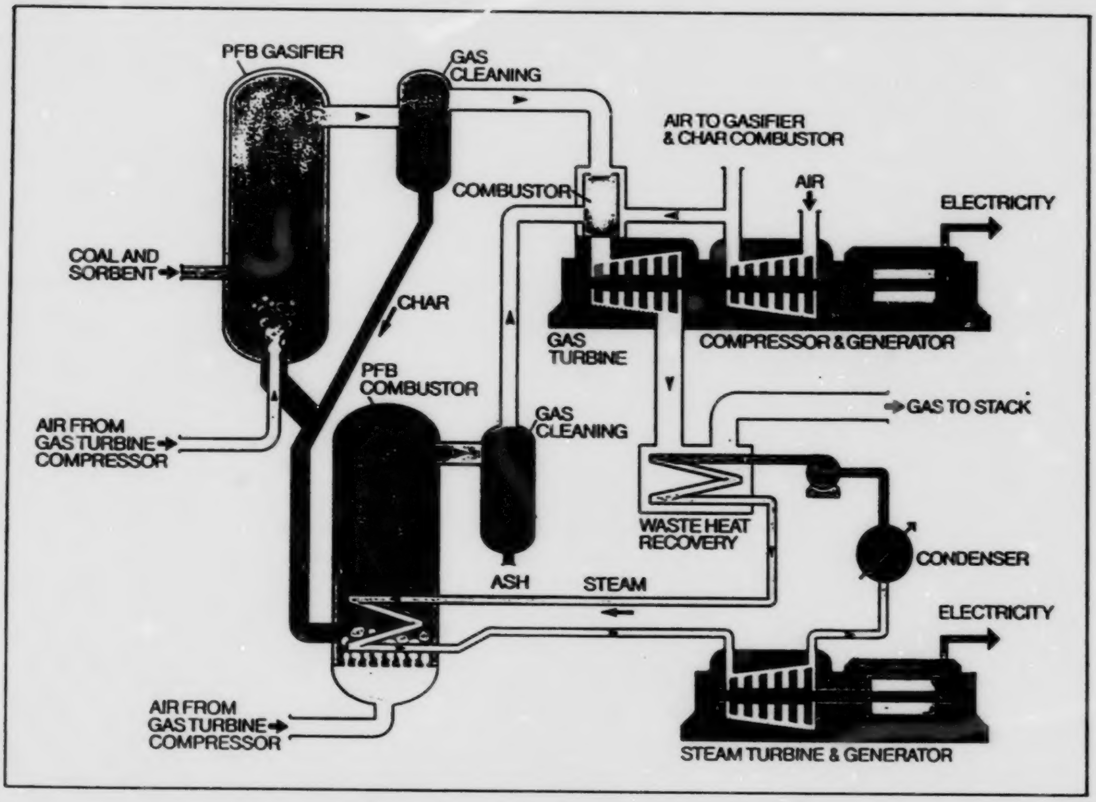

PFB GASIFIER
GAS CLEANING
AIR TO GASIFIER & CHAR COMBUSTOR
COMBUSTOR
AIR
ELECTRICITY
COAL AND SORBENT
CHAR
GAS TURBINE
COMPRESSOR & GENERATOR
PFB COMBUSTOR
GAS TO STACK
AIR FROM GAS TURBINE COMPRESSOR
GAS CLEANING
WASTE HEAT RECOVERY
CONDENSER
ASH
STEAM
ELECTRICITY
AIR FROM GAS TURBINE COMPRESSOR
STEAM TURBINE & GENERATOR

Figure 2.1

The British Coal Topping Cycle Based on a Pressurized Fluidized Bed Combustor

(Dawes)

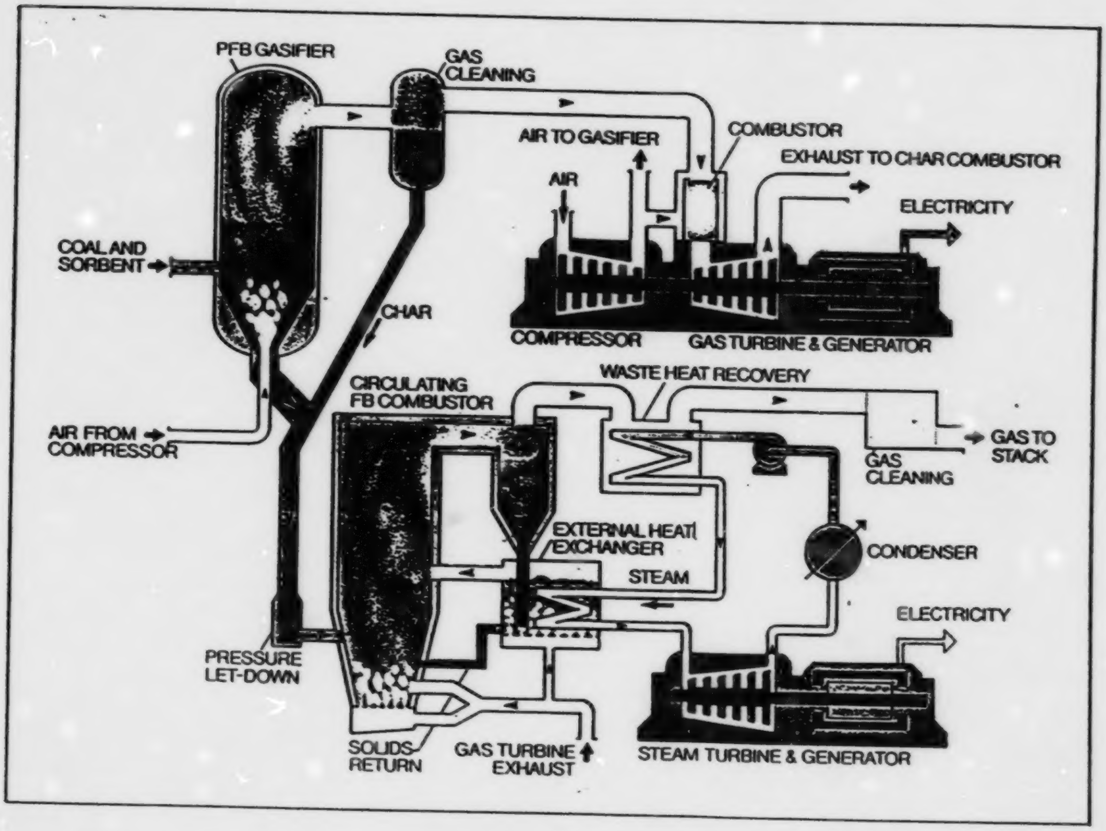

PFB GASIFIER
GAS CLEANING
AIR TO GASIFIER
COMBUSTOR
EXHAUST TO CHAR COMBUSTOR
AIR
ELECTRICITY
COAL AND SORBENT
CHAR
COMPRESSOR
GAS TURBINE & GENERATOR
WASTE HEAT RECOVERY
CIRCULATING FB COMBUSTOR
AIR FROM COMPRESSOR
GAS TO STACK
GAS CLEANING
EXTERNAL HEAT EXCHANGER
CONDENSER
STEAM
ELECTRICITY
PRESSURE LET-DOWN
SOLIDS RETURN
GAS TURBINE EXHAUST
STEAM TURBINE & GENERATOR

Figure 2.2

The British Coal Topping Cycle Based on a Circulating Fluidized Bed Combustor

(Dawes)

M.W. Kellogg is utilizing information gained in the development of the KRW total gasification system and their heavy oil cracking units to develop what they call a Transport Reactor System (see Figure 2.3). Gasification takes place in an up-flow entrained reactor. Heat transfer is carried out in a companion fluidized bed. Pilot scale process development is ongoing.

Vereinigte Elektrizitätswerke Westfalen (VEW) has been operating a 10 t/h, 2 MPa, 1600°C down flow entrained gasifier for over six years (see Figure 2.4). Ground coal (< 90 µm) is fed to the gasifier along with preheated air. The cooled char from the gasifier is combusted in a conventional pulverized coal fired design boiler.

The reactor section is followed by a radiant cooler that lowers the gas temperature to 1,300 - 1,400°C. Steam at 18.2 MPa and 535°C is generated in the radiant cooler and sent to the steam turbine. Approximately 10% of the ash in the coal leaves the reactor as slag (< 1% carbon composition).

After further cooling in a convective cooler, the coarse fraction of the char is separated in a cyclone at 800°C. The cooling in the reactor results in the need for less cooling in the convective cooler section, which contributes to the higher efficiency. The present configuration sees the char further cooled to 200°C and then fed to a pulverized coal boiler where electricity is generated. The low temperature energy is used in preheat applications. The commercial scale operation would send the 800°C char to a circulating fluidized bed combustor.

Chlorine and fluorine are removed in a system that utilizes lime in a circulating fluidized bed system at 2 MPa and 300°C. The solid waste from the dry scrubbing system can be sent to the char combustion CFBC to further enhance the utilization of the sorbent capacity of the lime. It is also possible to treat the sorbent from the dry gas cleaning separately and to produce a dry salt.

Testing is presently under way on electrostatic precipitators (ESP) to remove particles from the gas stream for turbine protection. The ESP's are operated at 350°C and 2 MPa. The target particulate concentration is 10 mg/Nm^3 (N refers to 0°C and 101 kPa conditions). In a commercial operation VEW will utilize a particulate removal system that is optimum for the application (ESP, ceramic barrier filter, etc).

VEW has 760 MWe of generating capacity based on cogeneration in operation at the same site. Natural gas turbines (three) are exhausted into three conventional pulverized coal fired boilers. This information is valuable for the design of a coal fired gasifier combined cycle operation. The plant efficiency is reported to be 42% and the availability similar to a conventional hard coal-fired power plant.

2.2.3.2 Co-Products

Activity in the development of processes that are designed to produce co-products is primarily by British Coal in the UK and as sponsored by the US Department of Energy (USDOE) in the US.

The feasibility of using the char or a char/coal blend as a substitute boiler fuel will be somewhat dependent upon the performance of the char in the combustor. If the char were to be fed to the Point Aconi CFBC, the characteristics of the char would be far less critical than if the utilization were to be in a conventional pulverized fired coal type boiler.

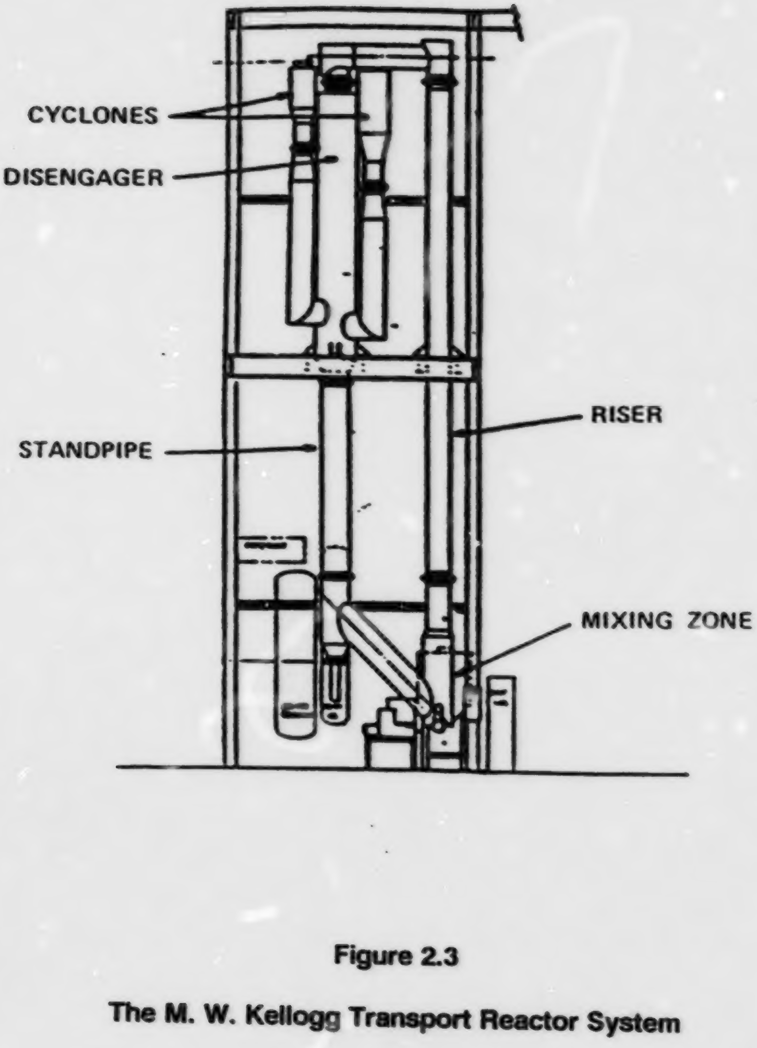

CYCLONES
DISENGAGER
RISER
STANDPIPE
MIXING ZONE

Figure 2.3

The M. W. Kellogg Transport Reactor System

(Banchik et al., 1989)

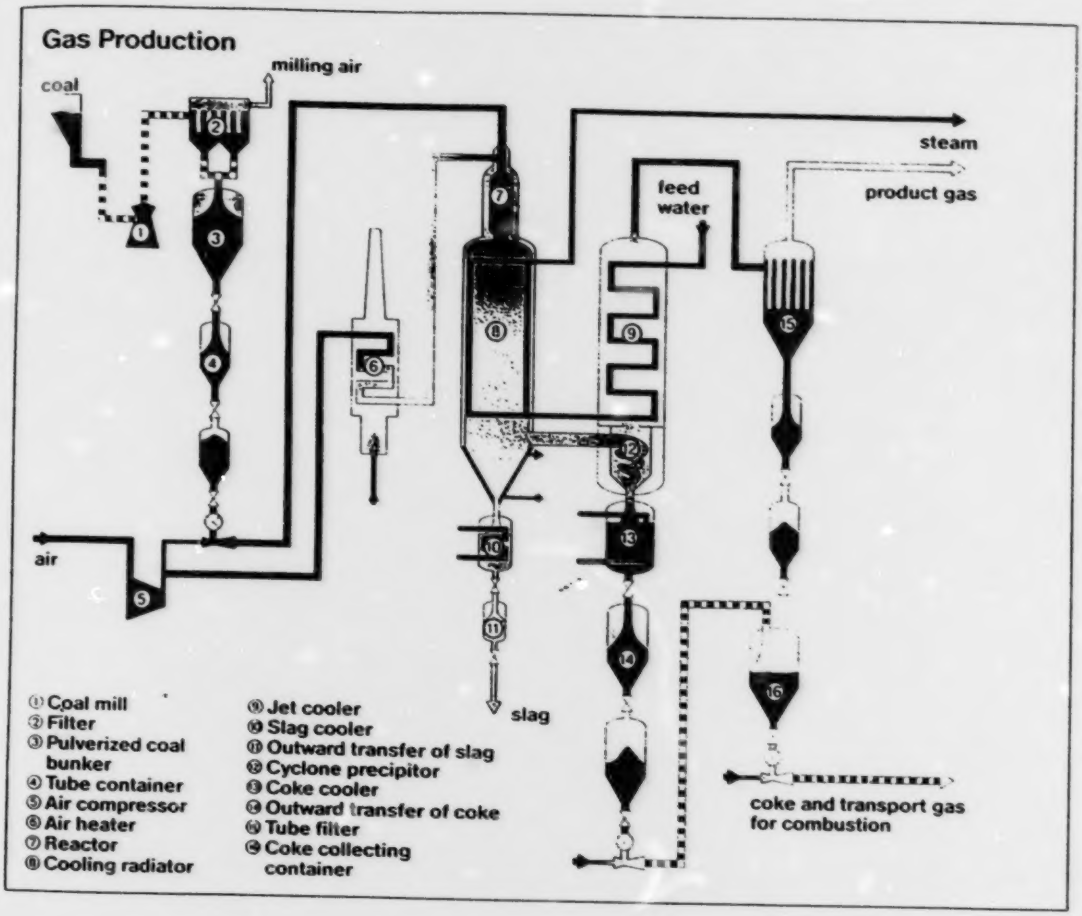

Gas Production
milling air
coal
steam
feed water
product gas
air
slag
coke and transport gas for combustion
① ② ③ ④ ⑤ ⑥ ⑦ ⑧ ⑨ ⑩ ⑪ ⑫ ⑬ ⑭ ⑮ ⑯
① Coal mill
② Filter
③ Pulverized coal bunker
④ Tube container
⑤ Air compressor
⑥ Air heater
⑦ Reactor
⑧ Cooling radiator
⑨ Jet cooler
⑩ Slag cooler
⑪ Outward transfer of slag
⑫ Cyclone precipitor
⑬ Coke cooler
⑭ Outward transfer of coke
⑮ Tube filter
⑯ Coke collecting container

Figure 2.4

A Schematic of the VEW Gasifier System (VEW)

British Coal

British Coal has developed two variants of a spouted bed gasifier. Both variants are designed to produce a gas product only. One variant utilizes a coal derived char bed while converting approximately 60 to 70%, mass basis, of the coal into gas. The other variant uses lower coal to air mass flow ratios and an inert bed (i.e. Bauxite) to convert up to 90% of the coal to gas. The majority of the fluidizing gas is introduced as a jet at the apex of the conical base. Additional fluidizing gas can be injected in the conical section to ensure complete fluidization. Solids circulation caused because of the gas flow prevents agglomeration of caking coal particles.

An atmospheric pressure, 0.5 tonne/h pilot plant was operated in both variants. The low conversion variant was operated for 700 hours. Bed material was approximately 80% carbon. Recycling of fines, that had been carried out (elutriated) of the bed with the gas flow, back to the bed resulted in an increase of only 2% (energy basis) in the conversion of coal to gas. Virtually no change was seen in the calorific value of the gas (3.1 MJ/m^3, net). Turndowns of up to 60% could be achieved. Up to 80% sulphur capture was achieved using limestone at a calcium to sulphur molar ratio of 2.

A total of 1750 hours of testing has been carried out using a high conversion variant of the gasifier. An inert bed material of silica sand or Bauxite was used to reduce the carbon content of the bed to between 5 and 10%. This higher density Bauxite bed results in the need for more fluidizing gas blower power. The use of silica sand as a bed material showed agglomeration which was as a result of interaction of iron oxide in the coal ash. Shallower bed depths of 1.5 m were used with the Bauxite bed which lowered the maximum coal conversion from 90% (1.5 m bed) to 86% (maf basis). The loss of conversion is more than offset by the reduction in the plant cost as the reactor size and weight is reduced. Turndowns to 45% of full load have been demonstrated with no reduction in the heating value of the gas.

Coal Technology Corporation Process

The design philosophy that the Coal Technology Corporation (CTC) has used is to convert coal to the highest-value-added products in a fashion that minimizes the plant capital costs in order to optimize the overall economics of the process. The co-products that have been targeted are char and coal liquids.

Data have been generated using three reactor systems to allow scale up to a process development unit (approximately 500 kg/h). The three units are a batch bench scale unit called the MGU, a 7.6 x 7.6 cm batch unit (BSR) used to simulate a block of coal in a 1 t/h unit and a 3.8 cm diameter single screw unit operated at Lawrence Livermore National Laboratory. In addition, liquid upgrading and utilization tests have been carried out.

In order to gain experience in the upgrading of coal derived liquids prior to generation of their own, CTC purchased 76,000 L of coal derived liquids from Coalite Ltd, Chesterfield, England. The Coalite process uses a vertical fixed bed reactor which is externally heated and has approximately 10 wt % yield of tar/oil.

CTC has purchased and erected a 160,000 L/d coal liquid refinery to upgrade the coal derived liquids. The refinery produces three fuel products simultaneously from coal liquids, diesel fuel, gasoline blending stock and heavy industrial fuel. CTC is marketing a blend of coal derived liquids and other fuels at nine service stations in southwestern Virginia.

The 454 kg/h (coal process rate) process development unit (PDU) has been constructed and is being tested.

Institute of Gas Technology (IGT)

The process that is being developed at IGT is a cooperative venture between IGT, Peabody Holding Company (representative of the largest coal producer in the USA), Bechtel National Inc. and Caterpillar, Inc. The process is intended to produce co-products consisting of solids and liquid chemicals. The liquid chemicals may be used to produce electrode pitch binder or chemical feedstocks. The solids would be used to produce formed coke for metallurgical applications, smokeless fuel for displacing firewood and activated char for gas phase contaminant removal.

Four major tasks have been undertaken:

Task 1	Literature survey of mild gasification processes, co-products upgrading and utilization, and market assessment.
Task 2	Bench-scale mild gasification study
Task 3	Bench-scale char upgrading study
Task 4	System integration studies

IGT has called their bench scale gasification unit a process research unit (PRU). The PRU has a fluidized bed (lower) section that is 20 cm ID and 2.43 m high. An entrained bed sits on top of the FB that is 10 cm ID and 3.96 m high. Heating is provided from electrical heaters that surround the bed and electrical heaters used to reheat the fluidization gas. Coal feed can be at the base of the FB section or in the freeboard section. Solids discharge is either from the bottom of the bed or in an overflow port 61 cm above the distributor plate. The conceptual commercial reactor would obtain its process heat from the recycle of hot char and/or by high-temperature flue gases derived from burning a portion of the coal derived gases.

Yield data can be broken down as follows (all yields are on a weight % of maf coal basis);

- 65% Char
- 30% Oils and Tars comprised of - 18% pitch, 4% heavy oil, 4% middle oil and 4% light oil
- 5% gas and water

A product development unit (PDU) has been designed to operate at 1 tonne/h. Funding from the USDOE has been approved to construct and operate the PDU.

The PDU will use an integrated char recycle system that blends the recycled char with the coal feed and thereby remediates the caking problems that were seen in the PRU unit. A recycled coal derived liquid will be used as a quench liquid in the spray tower. A thermal cracker will be added to allow processing of the pitch. Previous experience at IGT indicates that there should not be any problem with caking on the walls of an adiabatic unit.

2.2.4 Process Selection for More Detailed Description and Economic Evaluation

The most suitable process, in both the electrical generation and co-product formats, was selected for economic evaluation. The selection process put emphasis on: processes with proven performance and operability, suitability to handle Nova Scotia coals, flexibility to operate in a wide range of conditions and the associated environmental aspects of the process.

A weighted scoring procedure was used to allow an objective selection to be made. The process resulted in the selection of the Vereinigte Elektrizitätswerke Westfalen (VEW) process for electrical generation and the Institute of Gas Technology (IGT) process for the production of co-products.

2.2.4.1 Electrical Generation Process

An economic evaluation of a conceptual application of the VEW partial gasification process to the generation of electricity utilizing a Nova Scotia coal has been carried out. The conceptual generation scheme has been sited at the Nova Scotia Power Corporation's (NSPC) Point Aconi Circulating Fluidized Bed Combustion (CFBC) location. The VEW gasifier is combined with a CFBC to generate 284 MWe. NSPC cost data have been utilized in evaluating the CFBC related capital, operating and maintenance costs.

A single 2.65 m internal diameter gasifier operating at 2 MPa and 1600°C will accept 75 t/h of Prince coal. The gasifier will convert 60% (wt basis) to gas with the remaining going to the CFBC as hot char. The overall cycle efficiency (chemical energy, HHV, in coal to electrical energy to grid) is assumed to be 44%.

The coal sizing and handling system would be capable of handling coal (output) at 21 kg/s.

The air supply system to the gasifier would have a capacity of approximately 60 kg/s. It will be assumed that a stand alone air preheater will not be necessary as the air will be heated during the compression step.

Chlorine and fluorine will be removed from the hot gas prior to entering the gas turbine combustor. It is envisaged that a dry gas cleaning system will be used to optimize the system efficiency.

Approximately 8.4 kg/s of char at 800°C will be fed to the CFBC. The majority of the char will be collected in the cyclone separator after the convective cooler. Steam generated in the CFBC from the combustion of the hot char will contribute approximately 83 MWe to the plant output.

The radiant cooling section of the gasifier will produce approximately 86 kg/s steam at 18.2 MPa and 535°C. This steam will be combined with that produced from the combustion of the hot char in the CFBC and enter the steam turbine which is rated at 181 MWe. Additional heat recovery is carried out as the char is cooled further in the convective section to 800°C.

It is proposed that this conceptual system will be operated as a base load plant. This mode of operation will negate the necessity of operating two gas turbines. A single gas turbine rated at 103 MWe will be used.

Economic Evaluation of the Process

Detailed cost information has been generated by NSPC for the Point Aconi 165 MWe CFBC generating station. Some of this data has been utilized in this study to arrive at costs for the CFBC, steam generator and other associated equipment. The NSPC data indicate that a significant reduction in the capital costs are seen due to the utilization of common facilities. This advantage has been used in this analysis. In addition, information generated for NSPC by British Coal (Blamire, 1991) has been used to estimate capital costs.

The following procedure has been used to arrive at an estimate for the capital costs of the power plant.

The British Coal study has presented costs for a 350 MWe topping cycle system sited at Point Aconi. Data made available from NSPC for the Point Aconi CFBC has been used. The capital cost for the facilities that would be similar to the Aconi 165 MWe CFBC and used in the 284 MWe VEW gasifier/CFBC facility have been scaled using the following equation:

$$\frac{\$\ unit\ 1}{\$\ unit\ 2} = \left(\frac{MW1}{MW2}\right)^{0.7}$$

where \$1 and \$2 are the costs associated with units one and two respectively and MW1 and MW2 are the generation capacities of units 1 and 2 respectively.

All costs have been inflated at 5% to mid 1991 Canadian dollars.

The costs used have been scaled and inflated from the BC data unless otherwise indicated. The coal receiving, storage and reclaim facilities are available due to the development for the Point Aconi 165 MWe CFBC. The sorbent reception and storage is also available due to the development at Point Aconi.

The sand handling, combustor feed, feed handling, gasification, hot gas cleaning, gasification air compressor and gas turbine have all been adjusted to account for scale according to the equation given above.

The CFB combustor cost was scaled from the NSPC Point Aconi data.

The cooling water costs are at 20% of the BC amounts. This reflects the previous installation of the cooling water off-shore facilities, pumphouse, cooling water system intake and discharge and the outfall structure.

Electrical distribution costs are 20% of the BC costs due to previous development.

The rail spur costs were waived as they have been developed for the 165 MWe CFBC.

The infrastructure costs are at 20% of the BC values due to previous development.

A contingency of 14% has been added to reflect the lack of accuracy of the capital cost estimates.

Start-up Costs and Working Capital

Start-up costs have been estimated at one month's operation cost (excluding coal) + 25% of one month's fuel cost + 2% of the power plant capital cost.

Working capital has been determined to be equal to the cost of 60 days of coal, water, limestone, ash and spent sorbent disposal plus 1% of the capital requirement.

Operating and Maintenance Costs

The fixed operating costs have not been scaled from the 350 MWe BC costs. Linear scaling has been applied to the variable operating costs. Solid waste disposal costs have been calculated based on the coal feed rate and analysis.

Levelized cost data were used to arrive at the values shown in Table 2.6.

Table 2.6

Values Used in Determining the Cost of Electricity for the VEW Process

Values used to Determine the Cost of Electricity	
Power and Gasification Plant Cost	$1713/kWe
Start-up Costs	$38/kWe
CR (capital required)	$1751/kWe ($497M)
WC (working capital)	$40/kWe ($11.22M)
OMC (operating & maintenance costs)	$22M
FC (annual fuel costs)	$34.1M
EP (average annual power (net) production)	1990 GWh/y

The base case COE was determined using an 8% DCF rate of return. The discounted rate of return on investment was varied from 5 to 12% to show the effect on the COE. This effect is shown in the Figure 2.5.

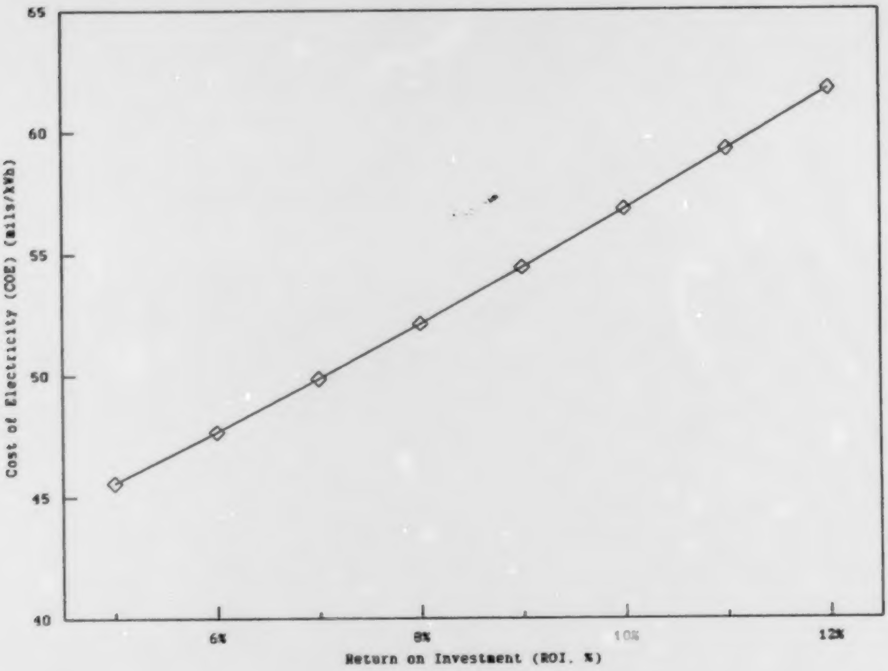


Figure 2.5

VEW Partial Gasification System, Sensitivity of the Levelized Cost of Electricity to the Return on Investment

The base case capital costs were varied by ± 40% to evaluate the COE sensitivity. The results are shown in Figure 2.6.

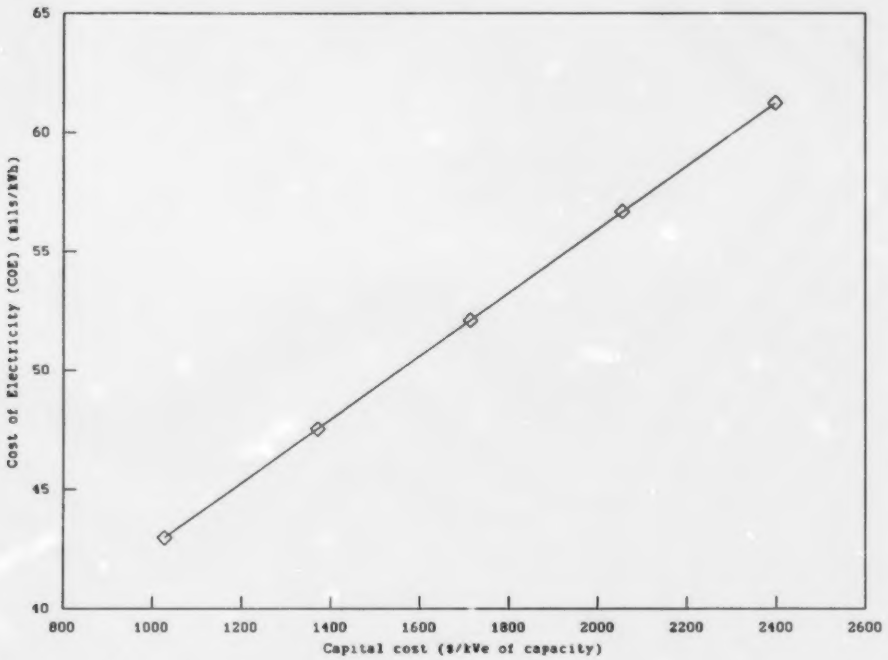


Figure 2.6

VEW Partial Gasification System, Sensitivity of the Levelized Cost of Electricity to the Capital Cost for the VEW System

2.2.4.2 Co-Product Process

The developers of the IGT process have projected the utilization of the char in three scenarios;

- char/coal blending for utility fuel
- metallurgical form coke and
- smokeless domestic fuel

Each of the scenarios is described below.

Char/Coal Utility Fuel

The char from the gasification process could be blended with thermal coal to produce a utility fuel. The local market for a blended thermal fuel of char and coal would be the Nova Scotia Power Corporation (NSPC).

If the char were to be fed to the Point Aconi CFBC, the characteristics of the char would be far less critical than if the utilization were to be in a conventional pulverized fired coal type boiler.

In the CFBC application, the sulphur can be fixed using a sorbent such as limestone, the higher ash content can be accommodated, variations in the ash fusion characteristics are less important due to the operation of the combustor at 850°C and the char particle size distribution may be acceptable as is from the gasifier system.

Form Coke

Form coke would be used in the reduction of iron ore in blast furnaces. Coke is no longer used in the making of steel in Nova Scotia.

Predictions are that despite the turndown in the production of pig iron in the US, the market for imported coke will increase due to the disproportionate decrease in the coke production capacity.

Metallurgical coke typically has sulphur at < 1%, ash < 9% and volatile matter < 1%. Calculated char characteristics (from Prince coal) indicate that the char would have sulphur at approximately 4%. This could be reduced by varying the feedstock and blending with a lower sulphur coking coal in the form coke step. Sulphur in the formed coke will generally be too high to be used in metallurgical processes. In addition, the presence of calcium reduces the reactivity of the coke. The ash in the char at 10% would need only a slight reduction. The char volatile matter is projected to be in the 9 - 16% range. This would be reduced in the form coke process.

Smokeless Fuel

Forming and pretreatment as carried out for the form coke use would be needed to produce smokeless fuel. Some additives may be needed to react with the sulphur and to ensure that the fuel maintains its form during combustion.

The target market would be the areas using anthracite, charcoal and firewood. Markets may exist in the US if legislation is introduced to restrict the emissions from domestic sources. Significant markets exist in Korea and other Asian countries.

Condensable Hydrocarbon Uses

These products have been described previously. Their potential uses are described below.

The light and medium oil fractions can be marketed to the chemical refining industry to make benzene, toluene, xylenes, phenol, cresols, xylenols, cresylic acids, naphthalene, indene and pyridine bases.

With severe refining to remove sulphur, nitrogen and oxygen, the light oil fractions could produce a BTX mixture suitable as a gasoline octane enhancer. Limitations on the aromatic content in domestic fuels may inhibit the application of these chemicals as octane enhancers.

Phenols are used in the production of fireproof phenolic foam insulation. Naphthalene is converted to phthalic anhydride which is used as a plasticizer.

The heavy and middle oil fractions consist of creosote oil that could be used as a blending feedstock for timber preservative. Environmental concerns may limit the use of this product. The heavier ends of the creosote oil fractions can be used to produce carbon black.

The medium-soft pitch can be used as binders in road and roof applications.

Thermal upgrading of the medium-soft pitch would produce a product that would be used to produce carbon electrodes which are used in the aluminum and steel industries.

Gas Utilization

Virtually all of the gas produced will be used on site to provide process heat.

Co-Product Process Economic Evaluation

The annual cash flow projections have been made for the three plant options. A plant book life of 30 years has been used. The return (interest rate paid) on debt and equity (ROI) financing has been assumed to be equal. As other costs have been manipulated, the ROI has been varied in an iterative calculation procedure so that the net present value of the sum of the net profit (after tax) is equal to zero. The net profit figure includes payments of all taxes.

A gross profit tax rate of 50% has been used. It has been assumed that losses in the gross profit figures may be cumulatively carried forward and subtracted from the gross profit gains. Income taxes are not paid until the cumulative sum of the gross profit is greater than zero.

The annual inflation rate used is 5% (including escalation).

The financing of the capital cost has been divided into three categories: equity at 45%, debt at 35%, government (grant or investment at 0% ROI) at 20%.

A salvage value of 10% of the original capital cost (inflated to a future value) has been added to income in the final year of operation.

The debt is repaid on a straight line declining balance basis over the book life. The return on the debt is at a fixed rate and in proportion to the outstanding debt amount.

The equity is repaid as a lump sum in the final year of operation.

The plant is assumed to be at 30% of full scale production in year three after the start of construction and 60% in year four and 100% in year five. The inflated credit values are weighted according to the percent of full scale operation.

Cost Sensitivity Analysis

A number of cost parameters were varied to evaluate their economic sensitivity. The effect of varying individual costs as the others are held at the "base case" are discussed below.

ROI Sensitivity to Capital Cost

The capital cost section of the analysis has the least accuracy. The base case data contains a contingency to compensate for the lack of accuracy valued at 40% of the plant capital costs. The results of the sensitivity of the ROI to the variations in the overall capital costs are shown in Figure 2.7. In all cases the ROI does not show a dramatic response to variations in the capital cost. This lack of sensitivity reduces the effect of the projected ± 40% accuracy of the study.

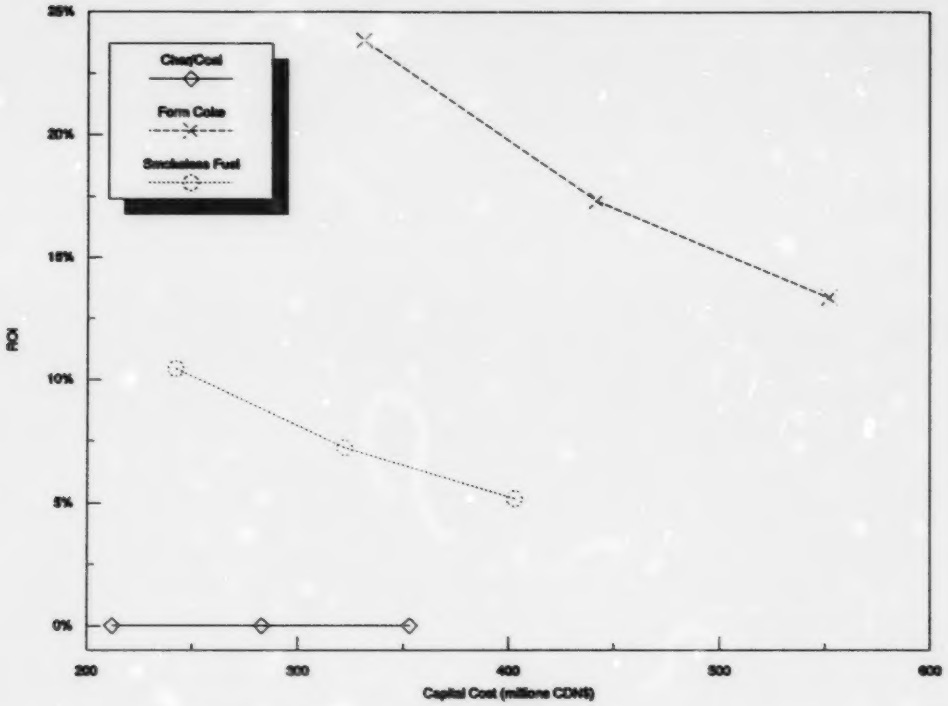


Figure 2.7

IGT Co-Product Process
Sensitivity of the Return on Investment to Capital Cost

The char/coal blend option does not show a positive ROI under any of the capital costs considered.

The form coke scenario projects the most favourable ROI as the capital costs are varied from the base case. ROI's vary from 24 to 13% as the capital cost is varied by -25% and +25% respectively.

The smokeless fuel option shows a marginal ROI (7%) under the base case situation. The ROI changes from 10% to 5% as the capital cost is varied by 25% (down and up respectively).

ROI Sensitivity to Coal Cost

Production costs result in Nova Scotia coal being relatively expensive (base case price of $65/t). Coal costs have been varied down from the base case in consideration that our coal is one of the most expensive. The resultant sensitivity analysis results are shown in Figure 2.8.

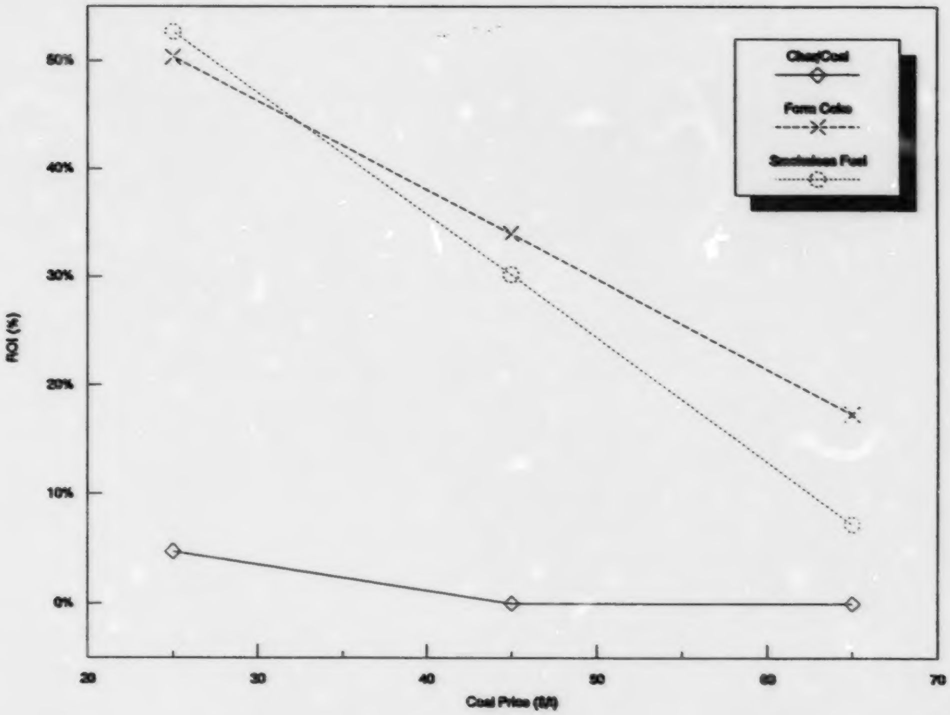


Figure 2.8

IGT Co-Product Process
Sensitivity of the Return on Investment to the Coal Price

The char/coal option shows little sensitivity to the coal cost and does not show a favourable ROI under any coal cost used.

Both the form coke and smokeless fuel options show significant sensitivity to variations in the coal price.

Sensitivity to Product Value

Figure 2.9 shows the ROI sensitivity to the value of the major product, char, form coke or smokeless fuel. The sensitivity of the ROI to product value decreases from the char/coal option to the smokeless fuel option and to the form coke option. Recent variations in the value of metallurgical coke increases the significance of the lower sensitivity to the product value.

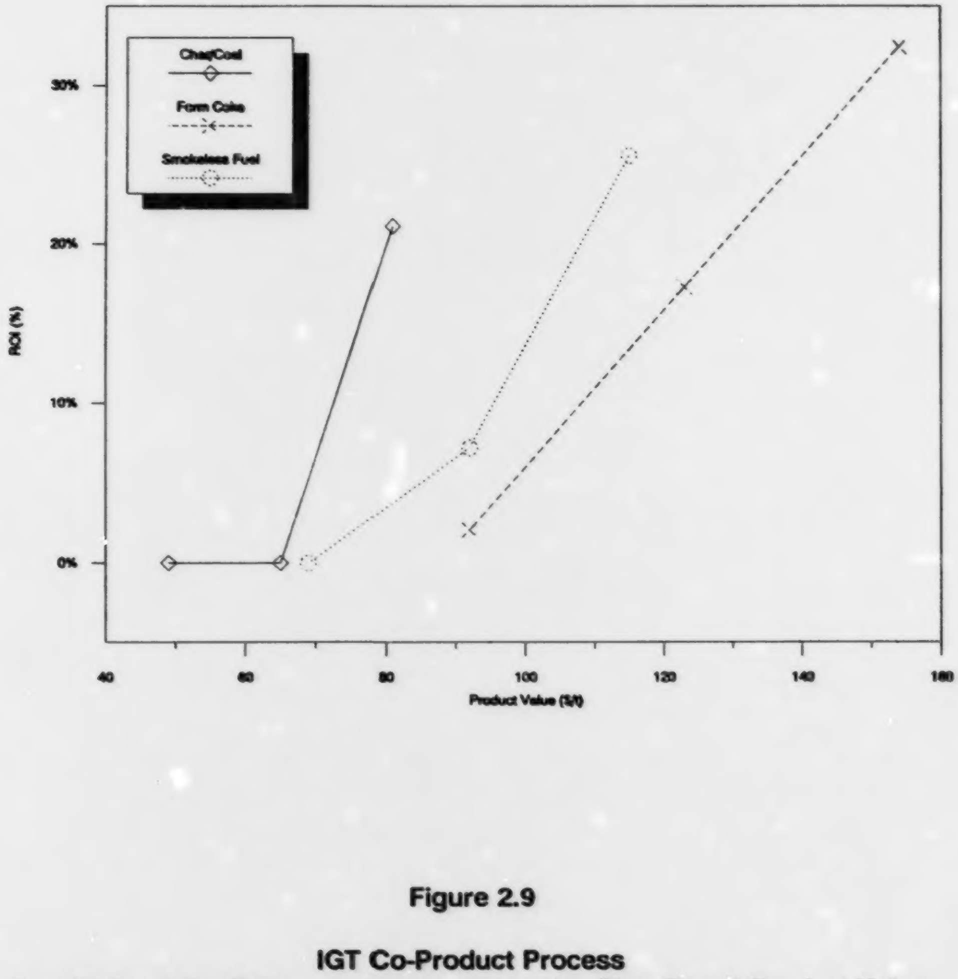


Figure 2.9

IGT Co-Product Process
Sensitivity of the Return on Investment to the Char Based Product Value

Figure 2.10 shows the relative value of the products for the three process options. The relative value will remain constant if the value of the products inflate at the same rate over the plant life. The sensitivity of the ROI to the char based product has been evaluated due to its dominant relative value in all three options.

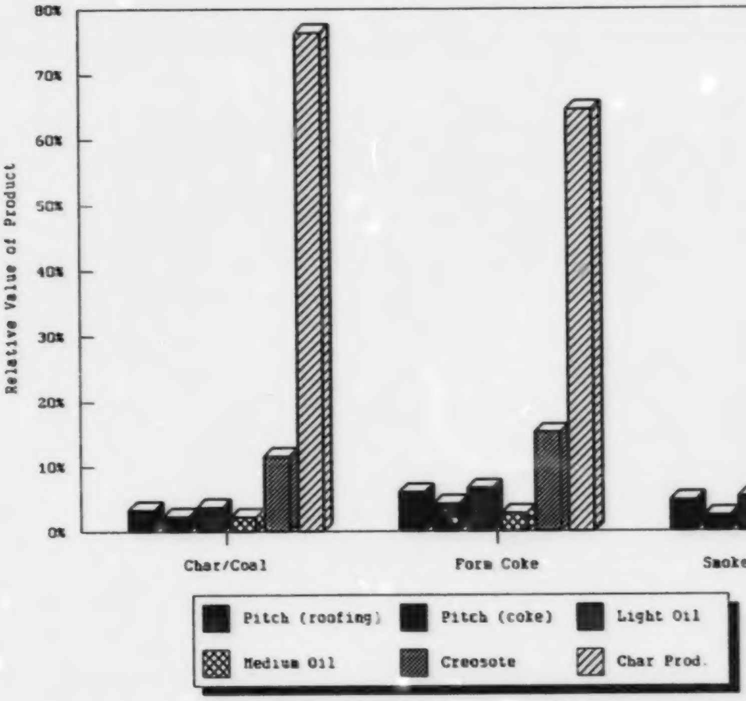


Figure 2.10

IGT Co-Product Process
Relative Value of the Products for the Three Process Options

3 CURRENT STATUS IN PARTIAL GASIFICATION OF COAL

3.1 Review of the Fundamentals Through to Bench Scale Partial Gasification

3.1.1 Coal Properties that Affect Gasification

This section of the report discusses some of the coal properties that can have an effect on the gasification process and the products that result. These effects are of greater importance under less severe gasification conditions. As the temperature and residence time of the gasification increase, the coal properties interact less with the reactor conditions and affect the product slate to a lesser degree.

Hamilton (1981) examined the effect of rank and heating rate on the development of plasticity during the pyrolysis of several Australian vitrinites. The greatest plasticity developed in bituminous ranked vitrinites. Cenospheres began to form in the vitrinites for bituminous ranks at heating rates of 10^{-1}°C/s.

Jenkins et al. (1973) used a Thermo-Gravimetric Analyzer (TGA) to prepare and examine the reactivity of chars produced from twenty-one American coals ranging in rank from lignite to anthracite. The chars were heated in a stream of dry N_2 at 10°C/min to 1000°C and held at that temperature for 2 hours or until weight loss stopped. The reactivity of the chars fell within a well defined band that showed a decrease with increased coal rank (or % carbon, wt % maf).

Furimsky et al. (1984) showed a strong correlation between mean reflectance (MR) and volatile matter (maf) for twelve Canadian coals (see Figure 3.1). The proximate and ultimate analysis of the coals studied are shown in Table 3.1. For bituminous coals, a positive linear relationship is shown between the H/C ratio and the yields of tars plus gases or the tar yields at 535°C in a Fischer assay retort (see Figure 3.2). A negative linear relationship is shown between the tar yields and the ratio of aromatic to total carbon (f_a) (see Figure 3.3). A correlation was also seen between the f_a of the tars and the f_a of the parent coals. Discrepancies between these results and those of others are attributed to the variation in the maceral and chemical composition of the coals.

Jenkins et al. (1973) also found that increased concentrations of magnesium and calcium in the char resulted in increased char reactivity.

Pyritic sulphur may physically be removed in the cleaning process such as froth flotation if the coal size is reduced so as to liberate the sulphur from the coal. Physical removal of the pyritic sulphur particle is difficult in Nova Scotia coals due to the fine dispersion of the pyrite in the coal structure. Decomposition of the sulphur takes place as the coal is heated.

Lambert et al. (1980), examined the production of pyrrhotites by pyrite reduction. Theoretical and laboratory data were used to demonstrate the reduction of pyrite to pyrrhotite in reducing an environment of H_2 at temperatures < 700°C. Pyrrhotite compositions were shown as a function of temperature and the concentration ratio of H_2 and H_2S. Complete conversion to pyrrhotite does not occur at temperatures < 450°C.

Table 3.1

Properties of Coal Feedstocks Tested by Furimsky et al., 1984

coal[a]	Proximate					Ultimate						f_a	MR
	Ash	Moist	FC	VM	VM maf	C	H	S	N	O	H/C		
1. Canmore	9.2	2.7	75.7	12.4	14.0	80.5	3.6	0.8	1.6	1.6	0.55	0.86	2.12
2. McIntyre	8.2	2.1	71.0	18.6	20.7	81.2	4.2	0.6	1.2	2.5	0.62	0.77	1.49
3. Byron Creek	17.3	5.1	54.0	23.5	30.3	66.5	3.8	0.3	1.0	6.0	0.69	0.74	0.93
4. Sukunka	3.2	1.9	70.7	24.0	25.3	84.0	4.7	0.5	1.3	4.4	0.67	0.79	1.41
5. Balmer	10.3	0.7	68.8	20.2	22.7	80.2	4.4	0.2	1.5	2.7	0.66	0.80	1.38
6. Coalspur	8.5	10.5	47.7	33.1	41.0	64.4	4.1	0.2	1.0	11.3	0.76	0.67	0.57
7. Shaughnessey	9.7	12.6	43.8	33.9	43.8	59.1	4.2	0.7	1.5	12.1	0.85	0.65	0.50
8. Devco	2.7	1.9	58.6	36.8	38.6	83.3	5.8	1.7	1.8	5.3	0.83	0.69	0.93
9. Prince	14.7	6.4	45.1	33.9	42.9	63.6	4.4	4.9	1.3	4.7	0.83	0.71	0.67
10. E. Blackfoot	21.2	15.4	30.1	33.3	52.0	44.9	3.7	0.3	1.0	13.5	0.99	0.58	0.49
11. Bienfait	7.0	31.2	32.0	30.0	48.4	45.0	3.0	0.5	0.8	12.5	0.80	0.57	0.40
12. Onakawana[b]	26.2	2.3	32.5	39.0	54.5	48.5	3.3	5.3	0.7	13.7	0.82	0.57	0.25

a. Bituminous coals No. 2 to 7 are from western Canada while No. 8 and 9 are from eastern Canada. Samples No. 11 and 12 are lignites.

b. Dried sample.

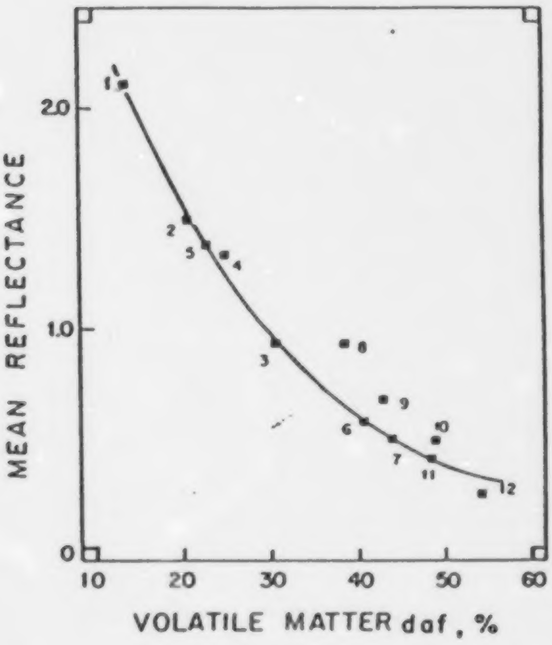


Figure 3.1

Mean Reflectance vs. Volatile Matter of the Coals Tested by Furimsky et al., 1984

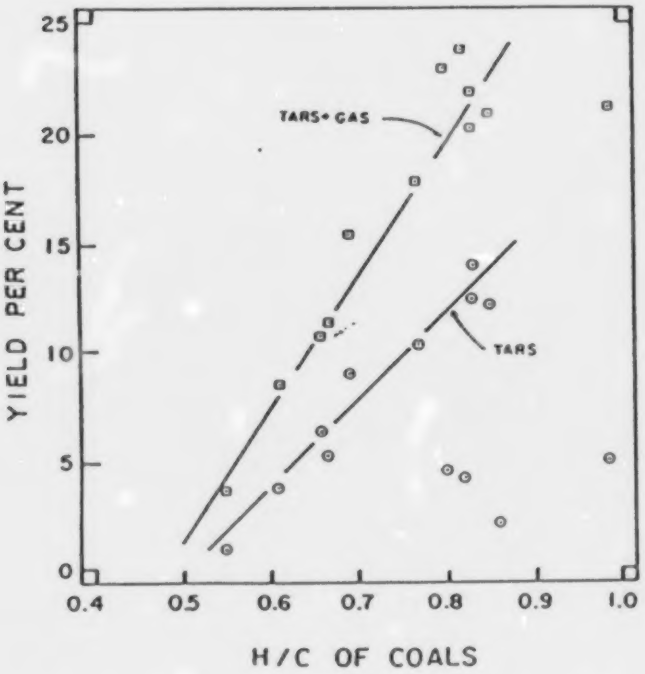


Figure 3.2

Yields of Tars and Gases vs. the H/C Ratios of the Coals Tested by Furimsky et al., 1984

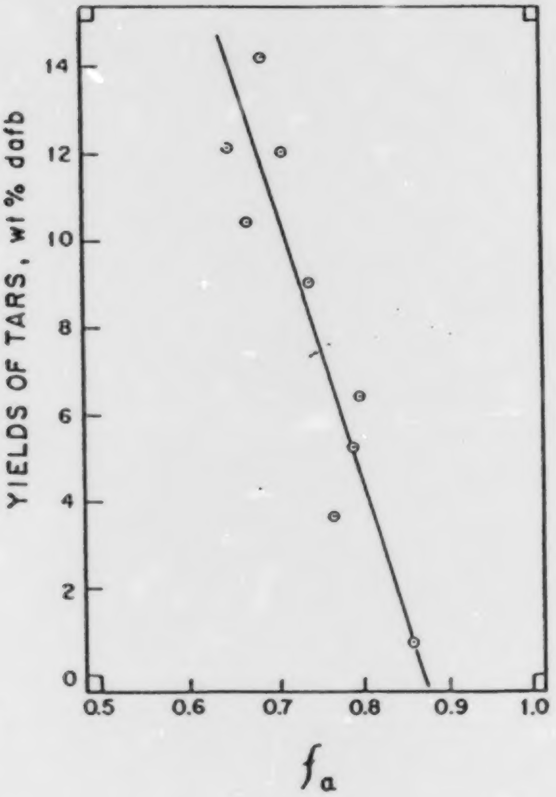


Figure 3.3

Yields of Tars vs. Aromaticity of the Coals Tested by Furimsky et al., 1984

Srinivasan et al. (1982), used electron spin resonance to evaluate the conversion of pyrite to pyrrhotite in atmospheres of vacuum, N_2 and H_2. Conversions began at 300°C in vacuum, 200°C in N_2 and 100°C in H_2. It is shown that differences in heating rate affected the temperature at which substantial rates of conversion began, between 300°C and 400°C. The strength of the C-H bonds in different coals will result in variations in the conversion in different coals under identical conditions.

3.1.2 Effect of Gasification Process Variables

Mahajan et al. (1980) carried out low level pre-oxidation of highly caking coals (HvAb, FSI = 8) in air at 120 - 250°C. The coals were pyrolysed in N_2 at temperatures up to 1000°C and produced chars whose reactivities were increased due to an increase in surface area.

Sotter et al. (1975) reported that the surface area of a char produced from a Wyoming sub-bituminous coal had increased almost two orders of magnitude from the parent coal. The char was made in the Occidental Research Corporation's entrained bed pyrolysis unit in which the heat source is from recirculated char.

Graff's et al. (1976) flash hydropyrolysis experiments resulted in yields of methane that increased steadily as the solids-contact time was increased from 2, 10 to 30 s. The entire yield of other hydrocarbons were obtained after 2 s.

Graff's et al. (1976) flash hydropyrolysis experiments showed that the molecular weight of the products decreased as the vapour phase residence time increased.

Stangeby et al. (1981) found that methane yield increased monotonically with peak temperature while the liquid yield optimum was at ≈ 800°C.

Jenkins et al. (1973) found that char reactivity increased up to temperatures of 600°C as a result of increased porosity. Further temperature increases resulted in decrease in the pore concentration and the reactive chemical structure both contributing to decreased char reactivity.

Singla et al. (1983) used a TGA to heat a HvBb coal to 150 or 200°C, held that temperature for 20 min and raised the temperature at 2.3 or 3.0°C/min and 8.2 or 8.3°C/min to 400, 600, 800 and 1000°C. Maximum macro and micro pore volumes were seen at 600°C with some collapse of the pore's structure after 800°C.

Vastola et al. (1981) has examined the effect of the coal particle size on the tar release rate under rapid heating gasification conditions. Coal particle sizes ranging from 53 - 1000 μm were tested at reactor temperatures of 500 to 900°C. It was found that the majority of the volatile release was completed within 300 - 700 ms and that even for the smaller coal particle sizes over 50 percent of the volatile release occurred under transient thermal conditions. This conclusion indicates the difficulty in assigning a gasification temperature when coal particles of various sizes are utilized.

Stangeby et al. (1981) evaluated the effect of heating rate (250 - 6000°C/s) on the yields from various Canadian coals including HvAb from Cape Breton, N.S. The heating rate was found to have little effect on the overall yield but to have dramatic effect on the composition of the products. Higher heating rates were found to increase the yield of lighter hydrocarbons.

Graff et al. (1976) carried out pyrolysis of Illinois No. 6 HvAb at 10 MPa in H_2 and N_2. At heating rates ranging from 20 - 1400°C/s, and in H_2 environment, the higher heating rates resulted in higher yields of the lower molecular weight products. It is discussed that higher temperatures and heating rates favour fragmentation over repolymerization.

Juntgen et al. (1982) showed in bench scale work using a high speed camera that the heating rate did not affect the coal swelling.

Block et al. (1975) evaluated the effectiveness of air, steam, N_2 and H_2 in the removal of sulphur on ten coals ranging in rank from HvAb, HvBb to HvCb. The swelling properties of the coals were reduced by preheating in O_2 at 300°C. The most effective condition for an air environment was 450°C. Higher temperatures resulted in excessive carbon losses. A temperature of 900°C was optimum for an N_2 environment. In a steam environment, a continuous reduction in S was seen up to 550°C. At higher temperatures, the loss of carbon became excessive. H_2 was not found to be effective below 850°C. The reduction in the potential recapture of the released S may be affected by using a combination of optimized gas/temperature combinations. It was noted that oxidative pretreatment had a beneficial effect on the hydro-desulphurization due to the retention and opening-up of micro pores in the coal that permitted ready transport of the gas through the coal.

Tipton (1980) describes the Occidental Research Corporation's "Convert-Remove" process for sulphur removal from char or coal. The convert step deals with lowering the organic sulphur component and the remove step affects only inorganic sulphur. The convert step is a hydrotreatment which produces H_2S which in turn is available to react with the in-situ sulphur scavengers to form inorganic sulphides. The remove step can be accomplished by a number of procedures. Occidental chose to use steam reduction to form H_2S.

Tipton (1980) notes that the Occidental Convert-Remove process produces better results if a series of short residence time cycles are used as opposed to one long residence time application. Total sulphur removals of 90% have been achieved

Juntgen et al. (1982) showed in bench scale work that the pressure had little effect on the total pore volume but that the pore diameter decreased with increasing pressure, resulting in an increase in the internal surface area. Pressure did not have any effect on the swelling of the coal particles. Tars produced under an N_2 environment had an increase in the H/C atomic ratio as the pressure is increased. The decrease was partially due to the change in the structure of the tar molecules produced. As the pressure of the N_2 was increased, the aromatic character of the tar decreased. An increase in the H/C ratio in the tar was seen under H_2 environment as the pressure was increased. Higher yields of tars are made by direct hydrogenation of the parent aromatics.

Gasifier reactor operational conditions can have a significant effect on the products produced under mild conditions (i.e. low temperature and short residence times). High temperature (i.e. 1500°C) conditions result in the production of low molecular weight species.

3.2 Potential of Nova Scotia Coals for Partial Gasification

3.2.1 Summary of Nova Scotia Coal Deposits With Potential for Partial Gasification

Nova Scotia has two areas where significant coal recovery operations are active, Cape Breton and Pictou County. Figure 3.4 shows the location of the mines.

Table 3.2 lists some of the data for the mines shown in Figure 3.4.

Table 3.2
Status of the Active Coal Mines in Nova Scotia

Mine/Seam	In Situ Reserves (10^6 tonnes)	Recoverable Reserves (10^6 tonnes)	Production (tonnes/day)
Prince/Hub	102	53	5,000
Lingan/Harbour	47	37	4,000
Phalen/Phalen	99	74	12,500
Donkin/Phalen	700	-	-
Westray/Foord	47	18	2,740

A new access tunnel is planned for the Prince mine. A conversion to retreat followed by closure of the mine is planned for Lingan. The Donkin/Harbour seam has not been developed to a working mine. Two main tunnels have been driven to the deeps. Further development is on hold indefinitely pending market driven needs. The Westray mine has started preliminary production.

Tables 3.3 to 3.6 show the analysis of the Cape Breton Development Corporation coals on an in situ, dilution free basis. It should be noted that the information in the tables is approximate. Variations in the analysis are seen depending upon the degree of cleaning the coal has undergone and from what part of the seam it was mined. Table 3.3 contains the proximate and ultimate analysis; Table 3.4 the sulphur related data and other miscellaneous data; Table 3.5 the petrographic analysis and Table 3.6 the ash data.

Tables 3.7 to 3.10 show the analysis of the Westray coal with the same breakdown as the CBDC tables described above. The petrographic analysis shown in Table 3.9 has been arrived at by averaging the core analysis data. The seam thickness data for each core sample was not used to weight the data resulting in the generation of representative values only.

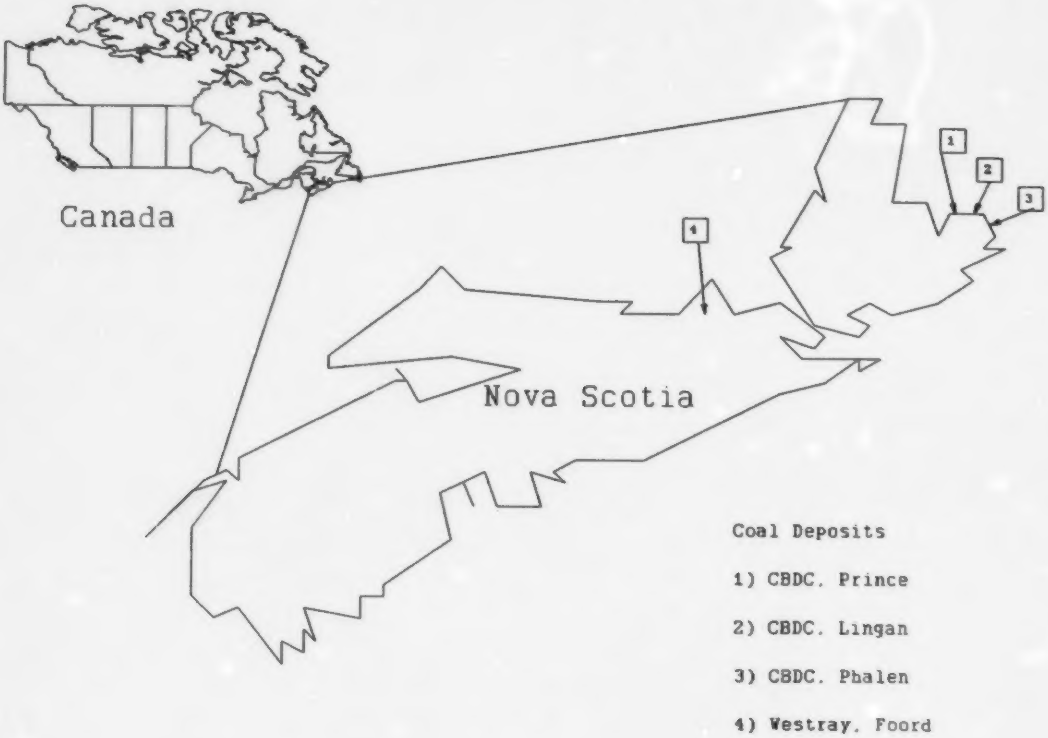

Canada
Nova Scotia
1
2
3
4
Coal Deposits
1) CBDC. Prince
2) CBDC. Lingan
3) CBDC. Phalen
4) Westray. Foord

Figure 3.4

Approximate Location of Coal Mines in Nova Scotia

Table 3.3

Ultimate and Proximate Analysis of the Cape Breton Development Corporation Nova Scotia Coal (In Situ, Dilution Free Basis)

	Mine/Seam		
	Prince/Hub	Lingan/Harbour	Phalen/Phalen
Proximate Analysis (dry basis)			
% Ash	7	5	8
% Volatile Matter	37	37	38
% Fixed Carbon	56	58	57
% Moisture	6	1.5	2.5
Ultimate Analysis			
% Carbon	76	81	77
% Hydrogen	5.4	5.2	5.1
% Nitrogen	1.4	1.7	1.5
% Sulphur	4	2	2.5
% Oxygen (by dif)	6.2	8.1	5.9

Table 3.4

Miscellaneous Analysis of the Cape Breton Development Corporation Nova Scotia Coal

	Mine/Seam		
	Prince/Hub	Lingan/Harbour	Phalen/Phalen
FSI	4	7	8
% Sulphur Total	4	2	2.5
Sulphur Forms			
% Pyritic	2.8	1.4	1.7
% Organic	1.2	0.6	0.8
H/C Ratio (molar)	0.85	0.77	0.79
$C_{aromatic}/C_{total}$ (molar)	0.71	-	-
Calorific Value, dry (MJ/kg)	31.4	33.9	32.6
Hardgrove Grindability Index	60	60	60

Table 3.5

Petrographic Analysis of the Cape Breton Development Corporation Nova Scotia Coal

	Mine/Seam		
	Prince/Hub	Lingan/Harbour	Phalen/Phalen
Vitrinite Types			
% V4	-	-	-
% V5	-	-	-
% V6	3.1	-	-
% V7	58.2	-	-
% V8	17.3	10.9	16.4
% V9	-	39.6	51.5
% V10	-	27.2	13.9
Reactives			
Total Vitrinite (%)	78.6	77.7	81.8
Reactive Semi Fusinite (%)	1.6	1.4	0.5
Exinite (%)	5.1	4.9	4.7
Total Reactives (%)	85.3	84.0	87.0
Inerts			
Inert Semi Fusinite (%)	3.3	2.7	0.9
Micrinite (%)	3.5	8.8	6.6
Fusinite (%)	4.0	2.5	2.9
Mineral Matter (%)	3.9	2.0	2.6
Total Inerts (%)	14.7	16.0	13.0

Table 3.6

Ash Analysis of the Cape Breton Development Corporation Nova Scotia Coal

	Mine/Seam		
	Prince/Hub	Lingan/Harbour	Phalen/Phalen
Elemental Analysis			
% SiO_2	27.6	28.3	34.1
% Al_2O_3	18.3	18.0	18.9
% Fe_2O_3	42.3	39.7	36.5
% TiO_2	0.9	0.8	0.7
% P_2O_5	0.08	0.3	0.4
% CaO	2.6	2.0	6.0
% MgO	0.9	1.2	0.9
% MnO	0.08	0.2	0.1
% Na_2O	1.2	0.8	0.8
% K_2O	0.9	0.6	0.9
Ash Fusion (°C)			
Oxidizing			
IT	1355	1365	1365
ST	1390	1400	1400
HT	1395	1415	1415
FT	1405	1445	1445
Reducing			
IT	1055	1075	1075
ST	1095	1110	1110
HT	1180	1160	1110
FT	1305	-	-
Trace Elements			
Cl	0.4%	0.2%	0.25%
Fl	24 ppm	20 ppm	-

Table 3.7

Ultimate and Proximate Analysis of Westray Nova Scotia Coal

	Mine/Seam		
	Westray/Foord		
	Run of Mine	Washed	
		Regular	Premium
Proximate Analysis (dry basis)			
% Ash	24.7	21.71	14.66
% Volatile Matter	26.0	28.24	30.24
% Fixed Carbon	49.3	50.05	55.10
% Moisture	-	8	8
Ultimate Analysis			
% Carbon	73.7	65.03	72.47
% Hydrogen	4.6	4.42	4.71
% Nitrogen	1.7	1.90	1.85
% Sulphur	0.8	0.74	0.69
% Oxygen (by dif)	5.1	6.20	5.62

Table 3.8

Miscellaneous Analysis of the Westray Nova Scotia Coal

	Mine/Seam
	Westray/Foord (run of mine)
FSI	5
% Sulphur Total	0.8
Sulphur Forms	-
% Pyritic	0.2
% Organic	0.6
H/C Ratio (molar)	0.75
$C_{aromatic}/C_{total}$ (molar)	-
Calorific Value, dry (MJ/kg) (maf)	30.6
Hardgrove Grindability Index	48 - 55

Table 3.9

Petrographic Analysis of the Westray Nova Scotia Coal

	Westray/Foord
Vitrinite Types	
% V4	NA
% V5	NA
% V6	NA
% V7	NA
% V8	NA
% V9	NA
% V10	NA
Reactives	
Total Vitrinite (%)	71.0
Reactive Semi Fusinite (%)	4.0
Exinite (%)	4.6
Total Reactives (%)	79.6
Inerts	
Inert Semi Fusinite (%)	0
Micrinite (%)	-
Fusinite (%)	6.6
Mineral Matter (%)	13.8
Total Inerts (%)	20.4

Table 3.10
Ash Analysis of the Westray Nova Scotia Coal

	Mine/Seam	
	Westray/Foord (washed)	
	Regular	Premium
Elemental Analysis		
% SiO_2	55.81	61.12
% Al_2O_3	22.39	23.06
% Fe_2O_3	9.00	6.91
% TiO_2	0.73	0.85
% P_2O_5	0.38	0.45
% CaO	3.12	1.98
% MgO	2.00	0.87
% SO_3	3.37	1.45
% Na_2O	0.65	0.82
% K_2O	2.55	2.49
Trace Elements		
Cl (ppm)	0.62	0.44
As (ppm)	56	34
Ash Fusion (°C)		
Oxidizing		
IT	1305	1298
ST	1457	1482
HT	1474	1482
FT	1482	1482
Reducing		
IT	1222	1228
ST	1372	1428
HT	1402	1462
FT	1427	1482

3.2.2 Relation of Nova Scotia Coal Deposit Characteristics to Partial Gasification Potential

The coal properties that are of significance are dependant upon the process conditions and design of the reactor. For example, the Vereinigte Elektrizitätswerke Westfalen (VEW) entrained bed reactor system would be rather insensitive to many of the coal properties such as free swelling index (FSI) or caking, atomic hydrogen to carbon ratio (H/C) and the ratio of aromatic to total carbon (f). $_a$In this case, the severity of the reactor conditions (1600°C and 2 MPa) and the tolerance to a variation in the gas phase products would make any variation in the coal characteristics less significant. The systems designed to produce co-products would generally be more sensitive to the properties noted above and others such as ash content, sulphur amount, FSI and petrographic analysis. The caking properties of Nova Scotia coals require special attention when a reactor design is chosen.

The ash content of the CBDC coals is relatively low. That of the Westray coal is in the 15 to 24% range. The high ash content of the Westray coal would result in an increased cost in handling during the processing steps of a partial gasification process.

The lower sulphur content of the Westray coal would require less hydrotreating of the liquid products and a form coke product would not need the addition of a sulphur sorbent. The lower hydrogen concentrations in the Westray coal would probably result in the need of more hydrogen consumption during hydrotreating of liquid products.

The lower iron content in the ash of the Westray coal may affect the gas and liquid yield data as the catalytic effect of the iron is reduced.

The work of Furimsky et al. (1984) indicates that as the atomic H/C ratio increases (Figure 3.2) the yield of the oils/tars increases. These data would imply the following order of decreasing oil/tars yields; Prince (atomic H/C = 0.85), Phalen (H/C = 0.79), Lingan (H/C = 0.77) and Westray (H/C = 0.75). Data are available on the f (f_a=$_a$0.71) for the Prince coal only which does not allow a comparison with the other coals.

Testing of the coals under actual reactor conditions is the only reliable method to ascertain the yield data.

3.3 Review of the Product Development Unit (PDU) to Commercial Scale Partial Gasification Processes that are Capable of Processing Nova Scotia Coals

This section presents a summary of the PDU to commercial scale partial gasification processes that are capable of processing Nova Scotia coals. The characteristic of Nova Scotia coals that is of most importance during processing is the caking qualities. The sulphur content is of significance in determining additional process steps and its effect on the product. The summary is divided into the non-active and active processes.

3.3.1 Non-Active Processes

A description of the non-active processes is provided to give perspective to the partial gasification development. In addition, operational data from non-active processes are sometimes much more comprehensive than those which are available from processes that are still under development. These data may provide some insight into the performance of active processes where their data are not available. In the cases where more detailed information has been reviewed, the descriptions will be more in depth.

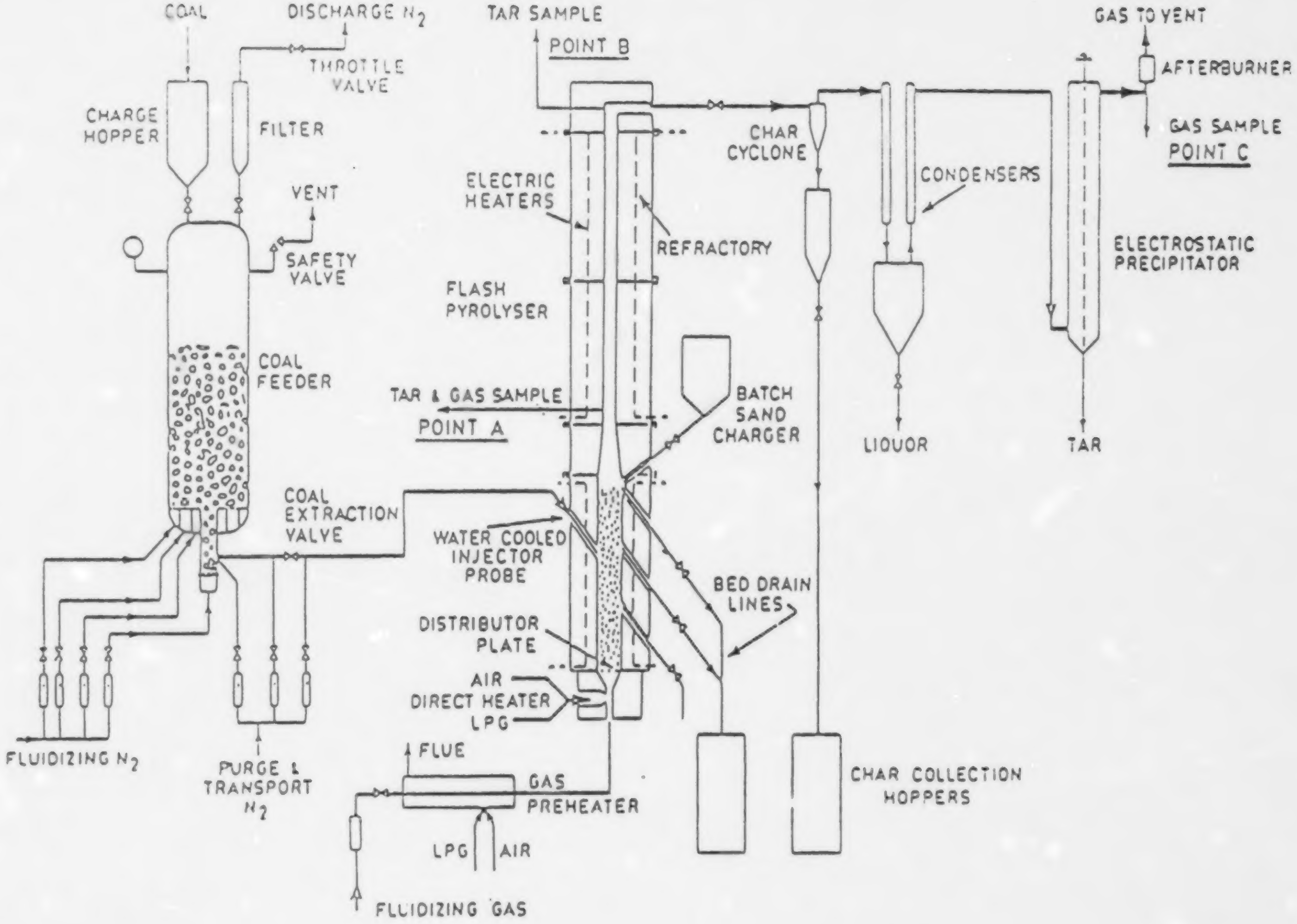
The CSIRO Pyrolysis Process

- Process Description

Operating and yield data have been generated from the CSIRO 20 kg/h PDU rig shown in Figure 3.5. Air dried coal sized < 0.178 mm was dense phase fed using N_2 as a transport gas through a water cooled injection probe to the FB pyrolyser. Heat for the pyrolysis was supplied by burning liquid petroleum gas (LPG) with a slight deficiency of air. Tar and gas yields were determined from samples taken at point A (Figure 3.5) just above the bed cross section reduction.

The desired heat source is hot solids circulated from a FBC. Introduction of the hot solids is from the base of the bed.

Char elutriated from the bed is collected in a cyclone. Figure 3.6 shows the oil scrubber/tar recovery system. The hot gas leaving the cyclone is quenched with an oil spray at two points followed by a venturi scrubber. A second cyclone separates the tar/oil mixture from the gas. Until sufficient pyrolysis oil is generated a coal-tar derived anthracene oil is used for scrubbing. A portion of the tar/oil mixture is cooled and recirculated through the quench system. Tar and oil mist that passes the second cyclone is separated from the gas stream in an electrostatic precipitator. The product tar is sent to a hydrogeneration unit to be upgraded to a crude feedstock.



COAL
DISCHARGE N_2
THROTTLE VALVE
CHARGE HOPPER
FILTER
VENT
SAFETY VALVE
COAL FEEDER
COAL EXTRACTION VALVE
FLUIDIZING N_2
PURGE & TRANSPORT N_2
TAR SAMPLE
POINT B
ELECTRIC HEATERS
REFRACTORY
FLASH PYROLYSER
TAR & GAS SAMPLE
POINT A
BATCH SAND CHARGER
WATER COOLED INJECTOR PROBE
BED DRAIN LINES
DISTRIBUTOR PLATE
AIR
DIRECT HEATER
LPG
FLUE
GAS PREHEATER
LPG
AIR
FLUIDIZING GAS
CHAR CYCLONE
CONDENSERS
LIQUOR
CHAR COLLECTION HOPPERS
GAS TO VENT
AFTERBURNER
GAS SAMPLE
POINT C
ELECTROSTATIC PRECIPITATOR
TAR

Figure 3.5

The CSIRO MK II 20 kg/h PDU

(Edwards et al., 1982)

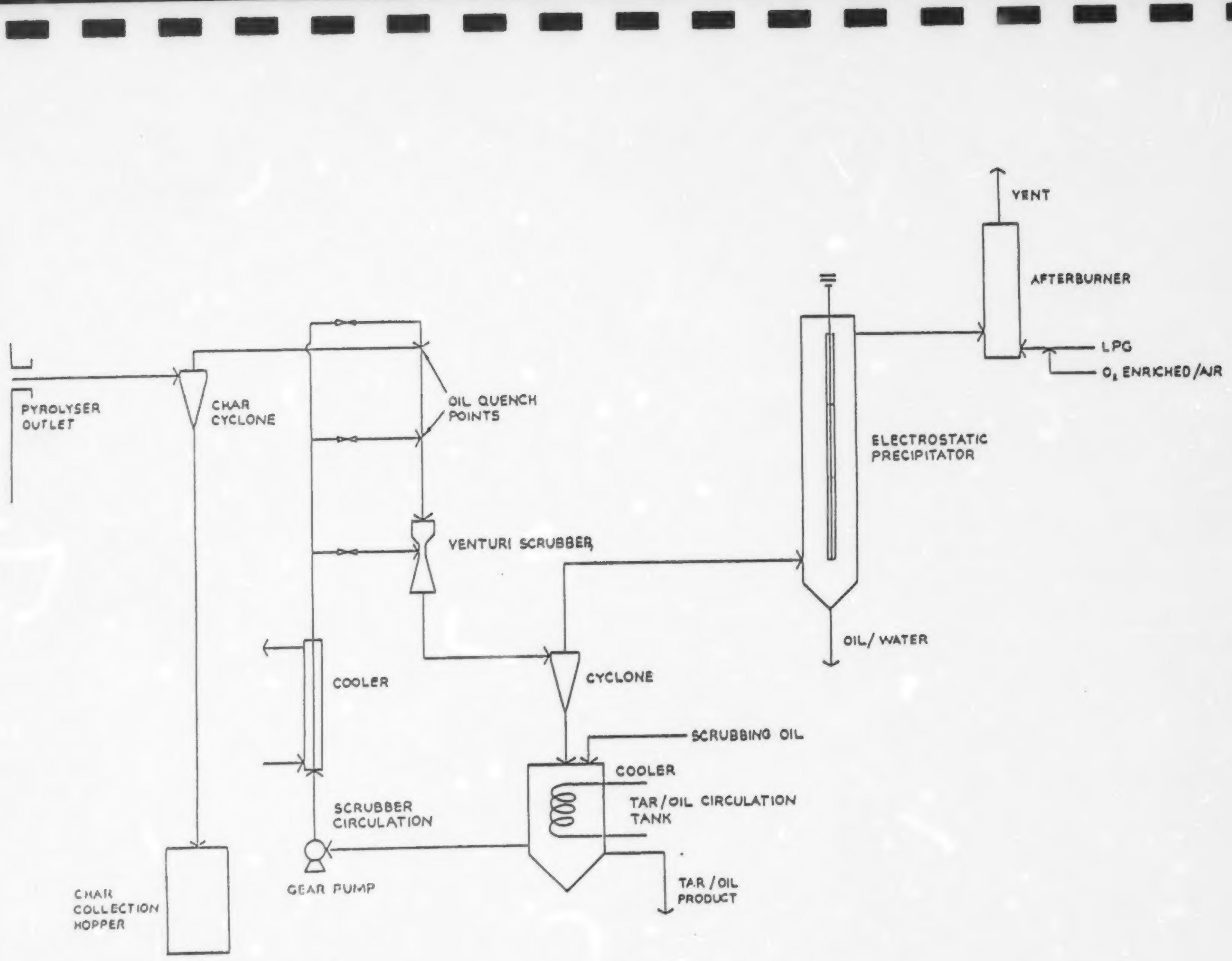

VENT
AFTERBURNER
LPG
O_2 ENRICHED/AIR
PYROLYSER OUTLET
CHAR CYCLONE
OIL QUENCH POINTS
ELECTROSTATIC PRECIPITATOR
VENTURI SCRUBBER
OIL/ WATER
COOLER
CYCLONE
SCRUBBING OIL
COOLER
TAR/OIL CIRCULATION TANK
SCRUBBER CIRCULATION
CHAR COLLECTION HOPPER
GEAR PUMP
TAR / OIL PRODUCT

Figure 3.6

Flow Diagram of Oil Scrubber/Tar Recovery System for the CSIRO 20 kg/h PDU

(Edwards et al., 1982)

- Yield Data

Operating data are shown in Table 3.11 for tests carried out with an Australian (Liddell) bituminous coal. The tests were conducted with the bed temperature between 550 and 690°C and fluidizing velocity in the FB section 3 to 6 times the minimum fluidizing velocity. Analysis of the Liddell coal is shown in Table 3.12.

Yield and performance data are shown in Table 3.13. The product yields were determined from samples taken just above the FB section (see Figure 3.5). CSIRO does not expect significant cracking of the tars to take place between the sample port location and the char cyclone. Due to the short duration of the individual tests, tar yields as measured by that collected in the recovery system (Figure 3.6) have been 60 to 80% of the sample port measurements. Quantities of tar are condensed on the surfaces of piping and not collected for measurement. Table 3.14 shows the analysis of the tar samples collected from the sample point. All samples show an increase in sulphur content from that of the parent coal.

The ash content of the char collected in the cyclone (Table 3.15) is nearly double that found in the coal. Short residence times and moderate temperatures result in volatile contents typically over 10% (air dried, wt %). It should be noted that 50 to 90% of the char remained as a coating on the bed sand. Longer duration tests may result in an equilibrium level of buildup. Only 10 to 15% of the char needs to be burnt in the envisaged parallel FBC. The char that is not burnt will be highly reactive and increase tar cracking when circulated to the FBC. This problem encountered when processing caking coals may require complete combustion of the char coating or severe preoxidation to eliminate caking. Preoxidation will also reduce the tar yields.

Figure 3.7 shows the proposed configuration of the heat generator and pyrolysis sections. As of mid 1982 the pyrolyser has been operated in conjunction with the tar collecting system and tar hydrogenerator. Char has been burned in the heat generator and hot solids circulation established. The char feeder is identical to the coal feeder. Batch feeders have been installed to allow the main feeders to be recharged on line.

The Occidental Research Corporation Process

Occidental's pyrolysis research was carried out on a bench scale unit (BSU) (2.3 kg/h) and a process development unit (PDU) (91 kg/h). The BSU was originally intended to get kinetic data and to aid in choosing operating ranges for the PDU operation. The BSU was later used to verify PDU runs and examine a low tar yield problem incurred in the PDU testing. The results of the PDU testing will be discussed below.

Table 3.11

Operating Data for the CSIRO 20 kg/h PDU Flash Pyrolysis Reactor - Liddell Coal

(Edwards et al., 1982)

RUN NO. DATE AVERAGE REACTOR TEMP. (°C)	L-1 12/12/77 615	L-4 12/4/78 610			L-6 26/4/78 690	L-7 2/5/78 700		L-8 9/5/78 550		L-11x 2/2/79 600	
Yield Sample No.	1	1	2	3	1	1	2	1	2	1	2
Reactor Temp. During Sampling (°C)	615	615	607	612	690	685	685	560	550	600	586
Reactor Pressure During Sample (kPa)	1.4	0.7	0.7	0.7	1.4	0.7	0.7	5.6	5.6	10.0	11.0
Fluidizing Gas Composition (% vol/vol) N_2 $CO + CO_2$ H_2O	80.0 8.6 11.4	79.9 8.6 11.5			75.2 10.6 14.2	78.3 9.3 12.4		83.7 7.0 9.3		85.3 6.3 8.4	
Gas Velocity in Fluid-Bed Section, V_{act} (m/s at reactor conditions)	1.43	1.43	1.42	1.43	1.58	1.71	1.71	1.58	1.56	1.53	1.50
Gas Velocity in Entrainment Section (m/s at reactor conditions)	8.22	8.24	8.17	8.21	9.11	9.88	9.88	9.09	8.98	8.83	8.61
Minimum Fluidizing Velocity, V_{mf} (1) Average sand particle size, dp = 0.68 mm (2) Average sand particle size, dp = 0.85 mm	0.258 0.403	0.260 0.407			0.243 0.380	0.241 0.377		0.271 0.423		0.259 0.405	
Ratio V_{act}/V_{mf} (1) dp = 0.68 mm (2) dp = 0.85 mm	5.54 3.55	5.50 3.51	5.46 3.49	5.50 3.51	6.50 4.16	7.10 4.54	6.50 4.16	5.84 3.74	5.76 3.69	5.91 3.78	5.79 3.70
Nominal Gas Residence Time (s) (1) Fluid-Bed Section (2) Entrainment Section (3) Between Coal Injector and Yield Sample Position	0.99 0.22 0.58	0.98 0.22 0.58	0.99 0.22 0.58	0.98 0.22 0.58	0.89 0.20 0.52	0.82 0.19 0.48	0.82 0.19 0.48	0.89 0.20 0.52	0.90 0.20 0.53	0.92 0.21 0.54	0.94 0.21 0.55

Table 3.12

Analysis of Liddell Coal Sample Tested in The CSIRO 20 kg/h PDU Flash Pyrolysis Reactor

(Edwards et al., 1982)

COAL SEAM COAL TYPE COAL LAB. NO.	LIDDELL/LIDDELL COLLIERY BITUMINOUS 59765
Proximate and Ultimate Analyses	
(1) Air-dried basis (% wt/wt)	
Moisture	3.32
Ash	28.08
Volatile Matter	31.25
Fixed Carbon	37.35
Specific Energy (MJ/kg)*	23.21
Carbon	55.32
Hydrogen	4.13
Nitrogen	1.41
Sulphur - (pyritic)	0.14
- (organic)	0.28
Oxygen (diff)	7.46
(2) Dry, ash-free basis (% wt/wt)	
Volatile Matter	44.55
Fixed Carbon	55.45
Specific Energy (MJ/kg)*	33.83
Carbon	80.64
Hydrogen	6.02
Nitrogen	2.06
Sulphur (organic)	0.40
Oxygen (diff)	10.88
Atomic H/C Ratio	0.896
Hardgrove Grindability Index (Volumetric Method)	51
B. S. Swelling No.	4.5
Size Analysis (% wt/wt as-fed)	
-185 + 152 μm	9.3
-152 + 106 μm	26.8
-106 + 75 μm	16.0
-75 + 53 μm	16.7
-53 + 45 μm	4.6
-45 μm	26.6

* Specific energy calculated by Dulong method and is given on an HHV basis.

Table 3.13

Yield and Performance Data for the CSIRO 20 kg/h PDU Flash Pyrolysis Reactor - Liddell Coal

(Edwards et al., 1982)

RUN NO. DATE COAL LAB NUMBER AVERAGE REACTOR TEMP. (°C)	L-1 12/12/77 59765 615	L-4 12/4/78 59765 610			L-6 26/4/78 59765 690	L-7 2/5/78 59765 700		L-8 9/5/78 59765 550		L-11x 2/2/79 59765 600	
Yield Sample No.	1	1	2	3	1	1	2	1	2	1	2
Reactor Temp. During Sampling (°C)	615	615	607	612	690	685	685	560	550	600	586
Coal Feedrate During Sampling (kg/h)											
(1) As-Fed Basis	11.8	15.5	14.4	14.8	15.2	12.0	13.2	19.8	19.0	18.0	19.0
(2) Dry, Ash-Free Basis	8.3	11.2	10.4	10.7	10.9	8.8	9.6	13.8	13.2	12.8	13.5
Gas Flowrate in Pyrolyzer (m^3/h @ NTP)											
(1) Wet Gas	29.2	29.1	29.1	29.1	29.9	32.3	32.3	35.8	35.8	34.6	34.6
(2) Dry Gas	25.8	25.7	25.7	25.7	25.6	28.3	28.3	32.5	32.5	31.7	31.7
Product Yields (% wt/wt maf coal)											
(1) Tar	29.0	29.8	29.0	24.6	23.0	27.0	24.6	31.0	29.0	30.3	32.3
(2) C_1 - C_3 Hydrocarbon Gases											
CH_4	2.20	2.67	-	-	-	-	-	2.10	2.10	2.20	2.50
C_2H_4	0.45	0.55	-	-	-	-	-	0.30	0.25	0.45	0.40
C_2H_6	0.67	0.89	-	-	-	-	-	0.80	0.72	0.72	0.80
C_3H_6	0.40	0.52	-	-	-	-	-	0.33	0.28	0.46	0.44
C_3H_8	0.26	0.35	-	-	-	-	-	0.36	0.32	0.30	0.32
Total C_1 - C_3	3.98	4.98	-	-	-	-	-	3.89	3.67	4.13	4.46
Coal Loading in Pyrolyser (kg maf/m^3 wet gas NTP)	0.286	0.383	0.401	0.366	0.374	0.274	0.297	0.387	0.369	0.370	0.390
Tar Loading in Pyrolyzer (kg/m^3 wet gas NTP)	0.083	0.114	0.103	0.090	0.086	0.074	0.073	0.120	0.107	0.112	0.126
Run Duration (hours)	1.4	1.9			0.7	2.5		1.4		2.0	
Total Coal Fed (kg)											
(1) As-Fed Basis	15.6	26.7			10.8	21.0		21.0		35.0	
(2) Dry, Ash-Free Basis	11.0	19.3			7.8	21.2		14.7		24.9	

Table 3.14

Analysis of Tar Samples Recovered During Yield Measurements on the CSIRO 20 kg/h PDU Flash Pyrolysis Reactor - Liddell Coal

(Edwards et al., 1982)

RUN NO. DATE	L-1 12/12/77	L-4 12/4/78		L-5	L-6 26/4/78	L-7 2/5/78		L-8 4/5/78		L-11x 2/2/78	
Yield Sample No.	1	1	2	1	1	1	2	1	2	1	2
Lab No.	60090	61107	61108	61149	61190	61191	61192	61222	61223	78713	78714
Reactor Temperature During Sampling (°C)	615	615	607	570	690	685	685	560	550	600	586
Tar Analysis (wt % maf)											
C	83.2	83.1	82.2	83.0	83.1	83.2	81.2	83.2	82.7	82.1	82.3
H	6.9	7.1	7.1	7.5	6.1	6.3	6.5	7.7	7.6	7.2	7.3
N	1.9	1.9	1.8	-	2.2	2.1	-	1.7	1.1	-	-
S	8.0	7.9	8.9	9.5	8.6	8.4	12.3	7.4	8.6	10.7	10.4
O (diff)											
Atomic H/C Ratio	0.995	1.025	1.037	1.084	0.881	0.909	0.961	1.110	1.103	1.052	1.064
HHV (MJ/kg) (Calc by Dulong Method)	36.6	36.9	36.4	37.3	35.3	35.7	34.8	37.8	37.3	36.3	36.6

Table 3.15

Analysis of Chars Produced in the CSIRO 20 kg/h PDU Flash Pyrolysis Reactor - Liddell Coal

(Edwards et al., 1982)

RUN NO.	L-1	L-4	L-7	L-8
CHAR LAB NO.	60091	61109	61205	61224
AVE. REACTOR TEMPERATURE (°C)	615	610	685	550
Proximate and Ultimate Analyses				
(1) Air-Dried Basis (% wt/wt)				
Moisture	2.7	1.6	1.3	0.8
Ash	51.7	50.9	48.9	45.2
Volatile Matter	11.6	11.8	9.9	16.9
Fixed Carbon	34.0	35.7	39.9	37.1
Specific Energy (MJ/kg)	14.3	15.0	15.4	17.01
Carbon	37.6	39.8	42.3	43.5
Hydrogen	1.7	1.6	1.4	2.4
Nitrogen	1.1	1.2	1.2	1.2
Sulphur	0.4	4.9	4.9	6.9
Oxygen (diff)	4.8			
(2) Dry, Ash-Free Basis (% wt/wt)				
Volatile Matter	25.4	24.8	19.9	31.3
Fixed Carbon	74.6	75.2	80.1	68.7
Specific Energy (MJ/kg)	31.3	31.5	31.0	31.5
Carbon	82.5	83.8	84.9	80.6
Hydrogen	3.7	3.4	2.8	4.5
Nitrogen	2.3	2.5	2.4	2.2
Sulphur	0.8	10.3	9.9	12.7
Oxygen (diff)	10.7			
Atomic H/C Ratio	0.538	0.486	0.396	0.670

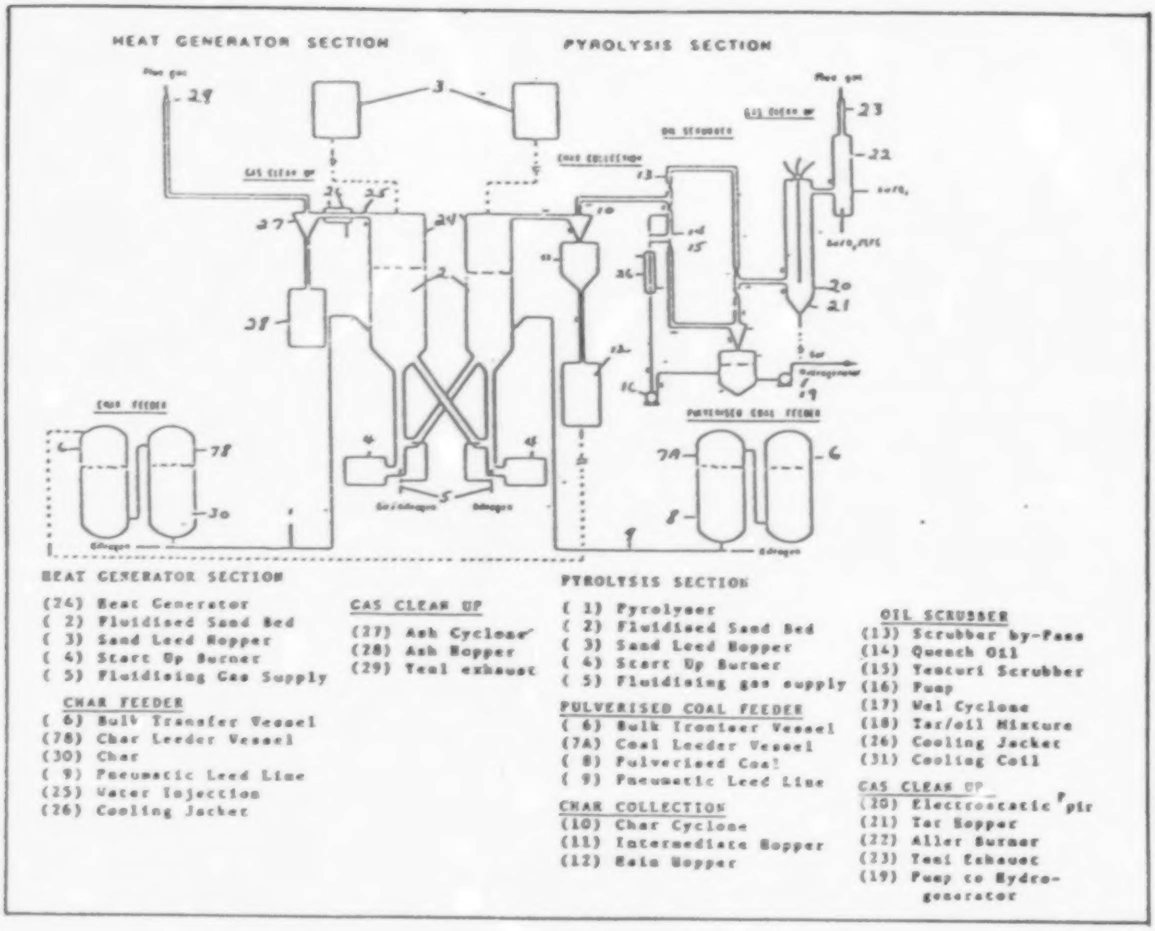

HEAT GENERATOR SECTION
PYROLYSIS SECTION
Flue gas
29
3
GAS CLEAN UP
CHAR COLLECTION
OIL SCRUBBER
GAS CLEAN UP
23
22
13
27
24
25
10
15
26
20
21
28
12
16
19
Tar
CHAR FEEDER
PULVERISED COAL FEEDER
4
4
5
Nitrogen
7B
30
7A
6
8
9
HEAT GENERATOR SECTION
(24) Heat Generator
(2) Fluidised Sand Bed
(3) Sand Feed Hopper
(4) Start Up Burner
(5) Fluidising Gas Supply
CHAR FEEDER
(6) Bulk Transfer Vessel
(78) Char Feeder Vessel
(30) Char
(9) Pneumatic Feed Line
(25) Water Injection
(26) Cooling Jacket
GAS CLEAN UP
(27) Ash Cyclone
(28) Ash Hopper
(29) Tail exhaust
PYROLYSIS SECTION
(1) Pyrolyser
(2) Fluidised Sand Bed
(3) Sand Feed Hopper
(4) Start Up Burner
(5) Fluidising gas supply
PULVERISED COAL FEEDER
(6) Bulk Transfer Vessel
(7A) Coal Feeder Vessel
(8) Pulverised Coal
(9) Pneumatic Feed Line
CHAR COLLECTION
(10) Char Cyclone
(11) Intermediate Hopper
(12) Main Hopper
OIL SCRUBBER
(13) Scrubber by-Pass
(14) Quench Oil
(15) Venturi Scrubber
(16) Pump
(17) Wet Cyclone
(18) Tar/oil Mixture
(26) Cooling Jacket
(31) Cooling Coil
GAS CLEAN UP
(20) Electrostatic Ppir
(21) Tar Hopper
(22) After Burner
(23) Tail Exhaust
(19) Pump to Hydro-generator

Figure 3.7

The CSIRO MK IIx PDU

(Edwards et al., 1982)

· Process Description

Figure 3.8 is a schematic of the Occidental PDU as it was first used. Crushed and dried coal was fed to the reactor using metered and heated N_2, CO_2 or a CO_2/H_2O mixture. Hot char, flowing at between 5 and 10 times the mass flow rate of the coal provides the heat for pyrolysis as it was mixed with the coal in the reactor. A series of three cyclones were used to separate the char from the pyrolysis vapours downstream of the reactor. The first two cyclones provided char for recycle to the reactor. Modifications to the char circulation system between these cyclones and the reactor will be discussed below. After passing the third cyclone the pyrolysis vapours went to the tar collecting system.

Figure 3.9 shows schematics of the original and Figure 3.10 the revised PDU reactor/char circulation system. The original char circulation system did not provide steady feed rate to the reactor (i.e. ± 20%). The revised system provided flows up to 1400 kg/h (± 5%) using pressure controls. The revised char circulation system as shown in Figure 3.10 saw the complete removal of the static bin and screw feeder. Momentary loss or reduction of char feed had resulted in the bituminous coal reaching the reactor wall before its caking qualities had been reduced and fouling of the reactor.

Figure 3.11 shows the actual reactor design. Near the end of testing the char distributing weir system in the reactor was modified to more evenly distribute the char and improve char-coal mixing.

The major area of modification and improvement to the Occidental system was in the char circulation system. The conclusion of the test program did not see the technical feasibility fully demonstrated for caking coal processing. The major problem was the physical layout of the system components which resulted in flow instabilities and char attrition. Figure 3.12 shows the proposed system configuration to alleviate the problems noted above.

Occidental has done bench scale research on removing the sulphur from the char produced from a high sulphur coal (Tipton, 1980). The Convert-Remove process is claimed to remove over 80% of the total sulphur and 50% of the nitrogen in a reducing environment.

The PDU tar collection system is shown in Figure 3.13. Raw pyrolysis gas enters the surge tank and then flows through a packed column and gas-liquid contactor counter current to a recycled quench liquid. Any entrained solids from the pyrolysis cyclones are removed with the heavy tars in this step. Part of the quench oil is recycled through a heat exchanger to the gas-liquid contactors. Gas leaving the contactors goes to a variable throat scrubber. Gas and non-condensed volatiles leaving the scrubber go to a venturi scrubber where light oils and pyrolytic water are removed and then separated in a decanter. The light oil is recycled to the surge tank and the water to the venturi scrubber. The gas goes to the demister and then is compressed and in this case sent to an incinerator.

Tar from the surge tank is separated into light and heavy fractions in a vacuum flash unit. The tar is preheated and flashed at 200 to 260°C, 0.7 kPa in a thin film evaporator. Some of the light tar is recycled to surge the tank. The gas separated from the light tar in the condenser is returned to the gas stream entering the demister.

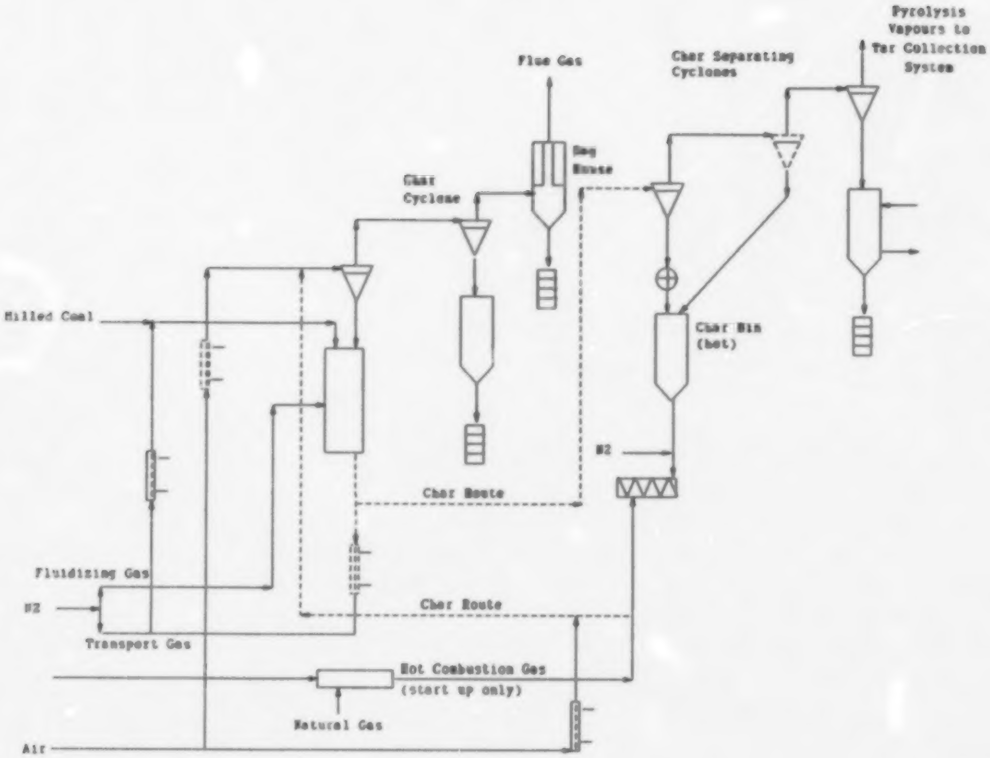
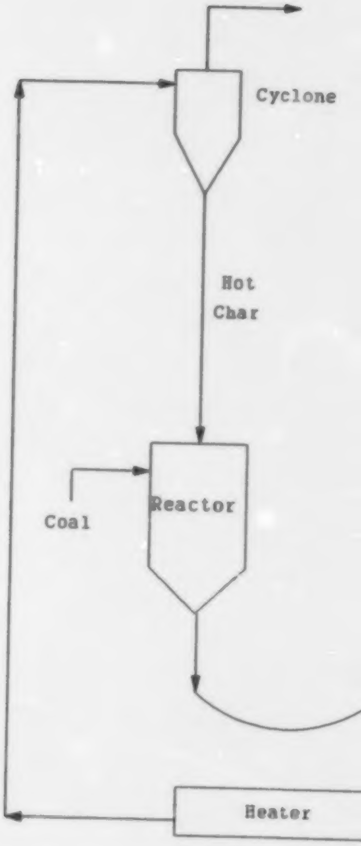

Pyrolysis
Vapours to
Tar Collection
System
Flue Gas
Char Separating
Cyclones
Bag
House
Char
Cyclone
Milled Coal
Char Bin
(hot)
N2
Char Route
Fluidizing Gas
N2
Char Route
Transport Gas
Hot Combustion Gas
(start up only)
Natural Gas
Air

Figure 3.8

The Occidental PDU as Originally Constructed

(Che et al., 1979)


Cyclone
Hot
Char
Coal
Reactor
Heater

Fig

The Occidental Research Char C
(Che e

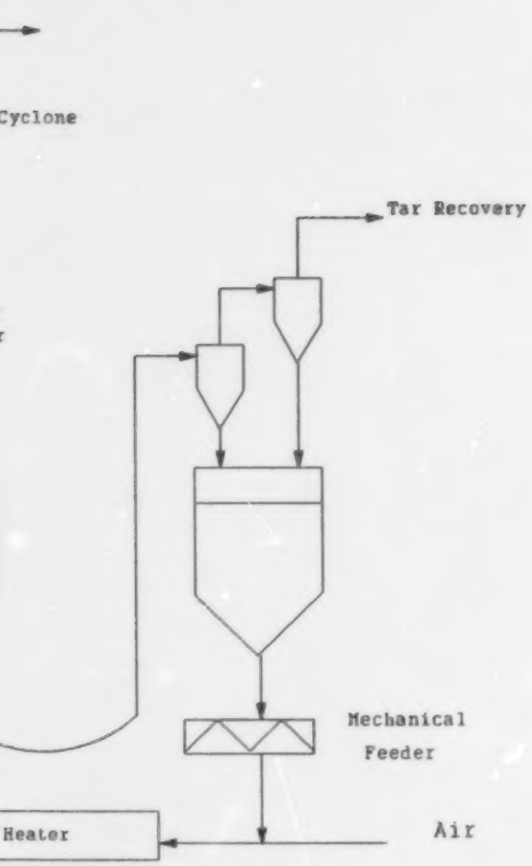

Cyclone
Tar Recovery
Mechanical
Feeder
Air
Heater

Figure 3.9

ch Char Circulation System in its Original Form
(Che et al., 1979)

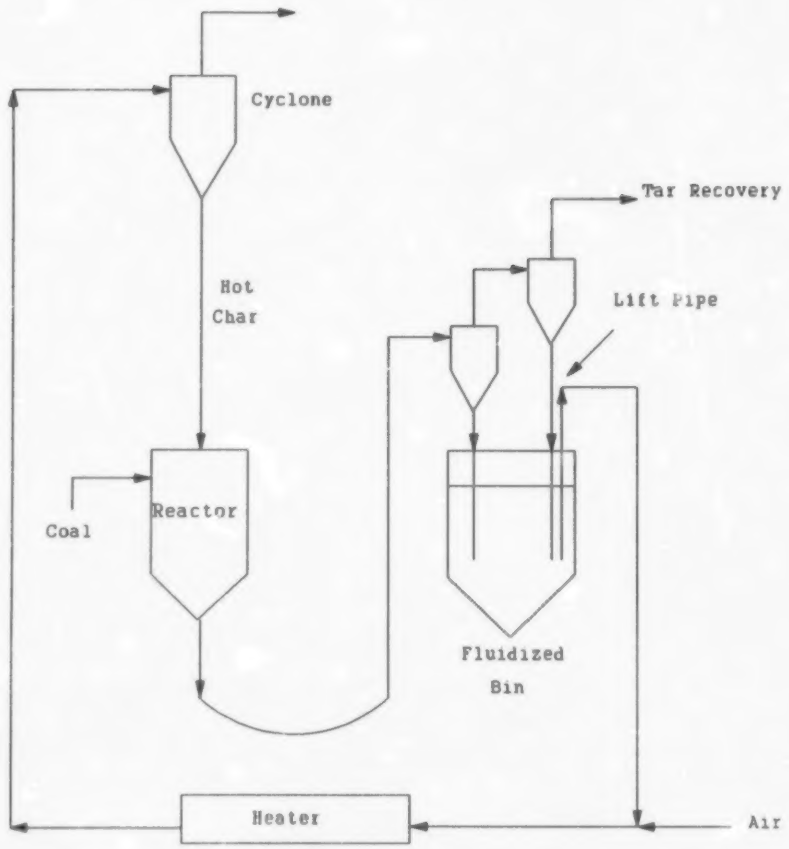

Cyclone
Tar Recovery
Hot
Char
Lift Pipe
Coal
Reactor
Fluidized
Bin
Heater
Air

Figure 3.10

The Occidental Research Char Circulation System in its Revised Form
(Che et al., 1979)

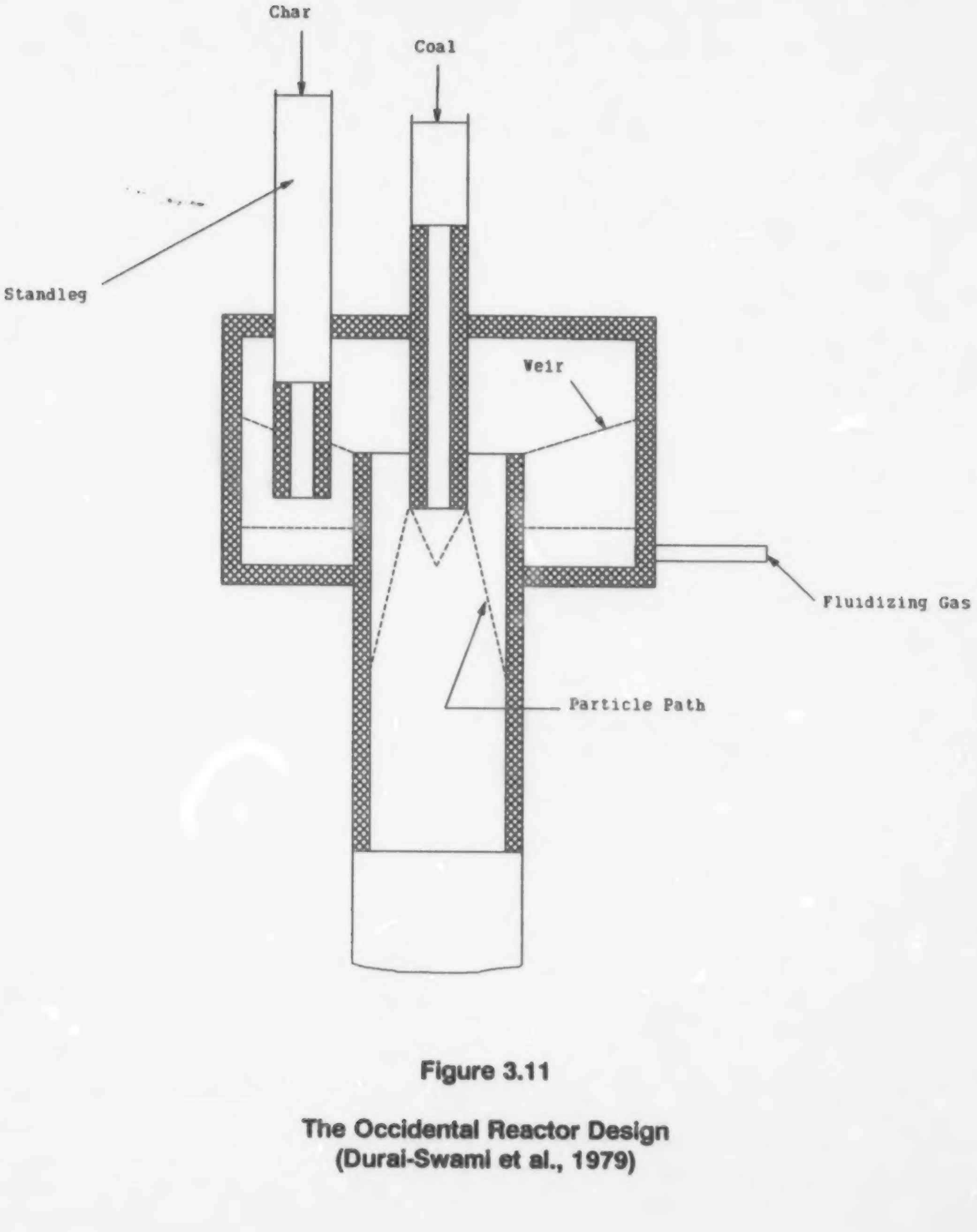

Char
Coal
Standleg
Weir
Fluidizing Gas
Particle Path

Figure 3.11

The Occidental Reactor Design (Durai-Swami et al., 1979)

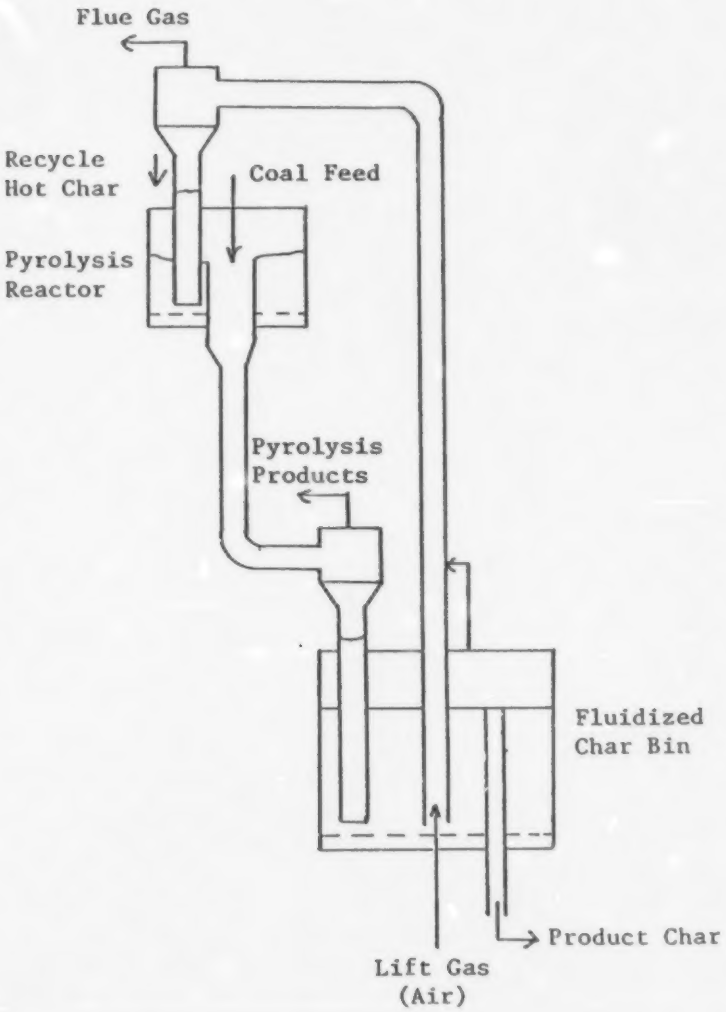

Flue Gas
Recycle Hot Char
Coal Feed
Pyrolysis Reactor
Pyrolysis Products
Fluidized Char Bin
Product Char
Lift Gas (Air)

Figure 3.12

The Occidental Proposed PDU System Configuration

(Che et al., 1979)

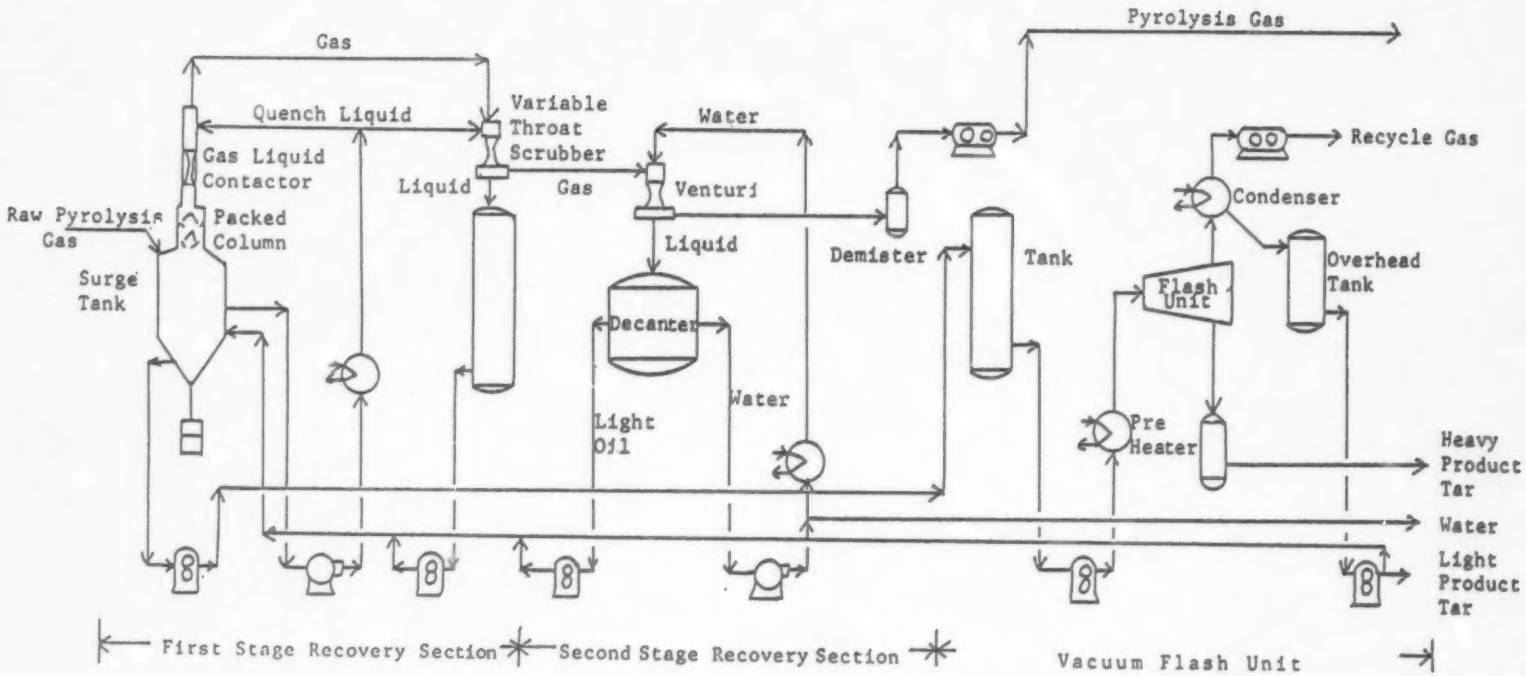

Gas
Quench Liquid
Variable
Throat
Scrubber
Water
Pyrolysis Gas
Recycle Gas
Gas Liquid
Contactor
Raw Pyrolysis
Gas
Packed
Column
Surge
Tank
Liquid
Gas
Venturi
Liquid
Demister
Tank
Condenser
Overhead
Tank
Flash
Unit
Decanter
Light
Oil
Water
Pre
Heater
Heavy
Product
Tar
Water
Light
Product
Tar
First Stage Recovery Section
Second Stage Recovery Section
Vacuum Flash Unit

Figure 3.13

The Occidental Tar Collection System

(Che et al., 1979)

- Yield Data

Tar yields and reactor conditions shown in Table 3.16 are for the pyrolysis of a West Kentucky # 9 bituminous coal (analysis shown in Table 3.17).

Tests 170 - 173 were made after shake down tests. The shake down tests were done using the sub-bituminous coals and resulted in modification to the reactor fluidization chamber and char circulation system. Tests 174 - 180 were done after the modified weir was added to the reactor to improve char distribution.

Due to the operating problems mentioned above, sufficient char was not produced to operate the system using recycled char. Fluid petroleum coke was used as a heat carrier. The surface area of the coke was only 17.5 m^2/g as opposed to over 200 m^2/g for a bituminous coal derived char. Both the increased surface area and reactivity of the char will lower tar yields. Reactive gases such as CO_2 and H_2O were used as carrier gases in some tests. It was expected that these gases would be absorbed on the surface of the coke and reduce tar cracking and polymerization. Tar yields for the bituminous coal tests were determined using a sampling train operating for periods of one to two hours. Samples were withdrawn directly downstream of reactor cyclones and at the midpoint of the reactor piping.
Referring to Table 3.18, the following conclusions may be drawn:

- Reactive gases such as CO_2 or H_2O reduce the char reactivity during pyrolysis and improve tar yields.
- Longer residence times in the reactor decrease the preasphaltene fraction of the tar yield.
- Optimum tar yields occur at 660°C and 1.3 second residence times.
- Tar yields greater than 200% of the Fischer Assay predicted tar content of the coal can be obtained.

The uncondensed light end fraction of two bituminous coal tests are shown in Table 3.18. Typically, this benzene rich stream is less than 2 wt % of the maf coal.

Table 3.19 shows the analysis of the product gases. The main components after removal of the added carrier gases are CO, H_2 and CH_4. Gas yields were not measured for the tests in which CO_2 was used as a carrier gas.

Modifications to the char heater/combustor to include secondary air injection improved the CO_2/CO ratio in the gas produced and reduced char consumption for the same heating effect.

Table 3.20 shows the yield and composition of the light oils (56 - 100°C BP) from a few PDU runs. As conditions are optimized the total tar yield increases, the light oil fraction increases and the atomic H/C ratio increases. Char yields for a bituminous coal are typically 43 wt % of the coal feed (Durai-Swamy et al., 1981).

Table 3.16

Yield and Properties of Bituminous Coal Tars for the Occidental Research Process

(Che et al., 1979)

Run Number	Heating Medium	Transport Gases	Reactor Temp (°C)	Residence Time Seconds	Tar Yield (Wt % maf)	Ultimate Analysis of Tar (Wt %)					Atomic H/C	Sp. Gravity (g/cc)	Solubility Classification (wt %)		
						C	H	N	S	O			Preasphaltenes	Asphaltenes	Oils
154	SB Char	CO_2	540	1.5	5.0	83.2	7.8	1.7	2.1	5.7	1.103	1.103	19	4	77
155	SB Char	CO_2	540	4.4	9.0	83.5	7.7	1.2	1.7	5.9	1.11	1.106	4	20	76
155	Fluid Coke	N_2	605	1.2	18.4	80.3	7.1	1.2	1.2	10.2	1.05	1.178	23	30	47
157	Fluid Coke	CO_2	605	3.8	28.0	81.2	6.9	1.8	1.7	8.4	1.02	1.155	12	29	59
158	Fluid Coke	CO_2	582	1.2	20.6	79.2	7.5	1.1	0.9	11.7	1.13	1.138	16	22	62
159	Fluid Coke	CO_2	605	1.4	30.3	77.4	7.4	1.2	1.5	12.5	1.14	1.114	19	22	59
160	Fluid Coke	CO_2	610	3.9	26.9	81.9	6.6	1.4	1.6	8.5	0.97	1.200	22	32	45
ESP Added															
170	Fluid Coke	CO_2	660	4.2	20.4	81.0	6.8	2.0	1.7	8.5	1.01	1.156	14	20	66
171	Fluid Coke	CO_2	605	1.5	25.9	79.2	6.9	1.8	1.7	10.4	1.05	1.146	16	24	60
171	Fluid Coke	CO_2 & H_2O	649	4.2	24.0	80.1	6.8	1.8	1.8	9.5	1.02	1.185	13	25	62
173	Fluid Coke	CO_2 & H_2O	649	2.1	27.2	80.9	6.6	1.9	1.0	8.8	0.97	1.214	28	22	50
174	Fluid Coke	N_2	660	3.8	24.3	81.0	6.6	1.8	2.0	8.6	0.97	1.174	13	26	61
175	Fluid Coke	N_2	660	1.4	26.0	78.8	7.1	1.8	1.7	10.6	1.08	1.143	18	21	61
176	Fluid Coke	CO_2	610	4.1	27.9	77.9	6.9	1.7	1.5	12.0	1.07	1.145	21	18	61
177	Fluid Coke	CO_2	610	1.5	34.6	80.0	7.1	1.5	1.0	9.6	1.06	1.152	14	29	57
178/179	Fluid Coke	CO_2	660	1.3	34.5	80.7	7.0	2.0	1.6	8.7	1.03	1.151	14	23	63
180	Fluid Coke	CO_2 & H_2O	643	1.3	31.9	82.1	6.4	1.7	1.6	8.2	0.94	1.207	35	7	57

Table 3.17

Analysis of the West Kentucky # 9 Seam Bituminous Coal Used in the Occidental Research PDU

(Che et al., 1979)

Proximate (wt %)		
Moisture	4.3	
Ash	8.3	
Volatile Matter	33.4	
Fixed Carbon	54.0	
	100.0	
Ultimate (Dry Basis) (wt %)		
Ash	8.7	
Carbon	72.3	
Hydrogen	5.2	
Nitrogen	1.6	
Sulfur	3.0	
Oxygen (by diff)	9.2	
	As Fed Basis	maf Basis
Fischer Assay (wt %)		
Tar & Light Oil	14.8	16.9
Water	10.1	6.6
Char	69.1	69.6
Gas & Loss	6.0	6.9
	100.0	100.0

Table 3.18

Composition of Uncondensed Light-End Fraction from Bituminous Coal Pyrolysis in the Occidental Research Process

(Che et al., 1979)

Run No.	179		180	
Uncondensed Light Oil (g)	56.6		28.7	
	wt % of Light Oil	wt % of maf Coal	wt % of Light Oil	wt % of maf Coal
Benzene	32.25	0.60	31.93	0.55
Toluene	3.93	0.07	10.41	0.18
Ethylbenzene	26.68	0.45	27.41	0.47
m- & p-Xylene	3.48	0.06	2.90	0.05
o-Xylene	1.30	0.02	1.62	0.03
Styrene	0.38	0.01	0.43	0.01
Naphthalene	0.02	trace	0.06	trace
Unidentified	28.96	0.49	25.24	0.44
Total	100.0	1.69	100.0	1.72

Table 3.19

Analysis of Product Gases from the Occidental Research Pyrolysis of Bituminous Coal

(Che et al., 1979)

Run No.	170	171	172	173	176	177
Temp. (°C)	660	660	649	649	670	610
Res. Time in Seconds	4.2	1.5	4.2	2.1	4.1	1.46
Gas Composition Volume % (Carrier Gas Free Basis)						
CO_2	C	C	C	C	C	C
CO	38.26	45.62	35.41	38.73	32.12	37.60
H_2	21.82	23.43	28.45	27.66	16.73	13.28
CH_4	25.57	13.70	23.63	20.96	33.68	27.25
C_2H_4	5.28	6.09	4.72	4.64	6.61	6.52
C_2H_6	3.32	4.19	2.88	2.99	5.54	5.67
C_{3+}	3.49	4.37	3.08	3.06	2.42	6.17
H_2S	2.26	2.60	1.83	1.96	2.90	3.51

C: CO_2 Used as Carrier Gas

Table 3.20

Yield and Composition of the Light Oil Fraction From a Bituminous Coal in the Occidental Research Pyrolysis PDU

(Che et al., 1979)

Run No.	172	173	180
Tar Yield (wt % maf coal)	20.5	24.8	29.4
Light Oil (wt % of tar)	4.5	4.1	12.3
Ultimate Analyses (wt %)			
C	83.2	80.9	76.1
H	8.4	9.0	9.1
N	0.4	0.5	0.3
S	0.7	1.0	0.9
O (by diff)	7.3	8.6	13.6
Atomic H/C	1.21	1.34	1.44
Identified Components wt of Light Oil			
Benzene	ND	ND	0.08
Toluene	0.28	0.12	3.72
p- & m-Xylene	1.19	0.78	0.58
o-Xylene	1.61	1.13	0.68
Phenol	7.37	4.70	2.70
Cresols	0.65	3.48	2.79
Xylenols	3.19	3.59	2.61
Naphthalene	3.84	3.50	7.32
Methylnaphthalene	2.38	1.70	1.65
Dimethylnaphthalene	0.47	0.27	0.47
Indene	3.69	3.18	0.07
Diphenyl	0.19	0.10	-

ND: Not Detected

- Further Development Needed

To summarize the results of the Occidental research the following further research is necessary:

- The final suggested system configuration should be tested (Figure 3.10) at the PDU scale.
- BSR studies should precede the above, with both using a candidate Nova Scotia coal.
- PDU scale testing using a bituminous coal generated recycled char should be done to properly determine the tar yields, gas yields and composition of both.
- The process should be carried out with lime or limestone present.
- The optimum temperature and residence time in the pyrolysis reactor for maximum liquid yields or maximum quality (i.e. H/C ratio, O, N, S content) need to be identified.
- PDU scale testing over a period sufficient to produce liquid yield data from actual liquid yields should be done.
- BSR and PDU testing would allow an evaluation of the stability (attrition) of the char from the high FSI Nova Scotia coals to be made.

The Lurgi-Ruhrgas (LR) Systems

Lurgi and Ruhrgas have combined efforts starting in the 1940's to design and construct a number of coal gasification facilities. The process was modified to its final version in 1951 in which the process heat is carried by product char. Facilities have been constructed ranging in size from bench scale to fully commercial (Rammler, 1981, 1982). Pilot plant operations at 10 t/d have been carried out in Dorsten, West Germany. Commercial plants operate in Yugoslavia, producing formed coke from lignite; Prosper coke-oven plant in Bottrop, West Germany; and in the Normandy Park Works of British Steel Corporation in Scunthorpe, England. The plants can be designed to operate at low temperatures and produce formed coke, low temperature and produce char and tar or at medium temperatures when coupled to a boiler.

A problem area in dealing with high swelling coals, as in Nova Scotia, is described in Rammler's 1982 paper. He notes that such coals may produce chars that are very friable and that break into fine particles. The fine particles of char would not be suitable as a circulating heat carrier. In addition, deposits of fines would build up on the condensation unit and other vessels and pipes. For strongly caking coals, higher temperatures will be needed and the process must be monitored closely to ensure that large agglomerates are not formed.

The pyrolysis process has been investigated mainly when operating at 590 and 900°C. At 590°C the liquid yields are optimized and at 900°C gas yields are increased. The same general equipment has also been developed as a hydrocracking unit using sand as the heat carrying medium. Table 3.21 summarizes the LR commercial and test facility installations.

A description of the equipment used for the low (590°C) and high (900°C) temperature carbonization at the commercial and test level is given below.

Table 3.21

A Summary of Lurgi-Ruhrgas Pyrolysis Plants

Operation	Location	Temperature	Application	Scale	Plant
Pyrolysis	Germany	900°C	Gasification	0.2 t/h	LR I
Pyrolysis	Germany	590°C	Maximum Tar Yield	0.2 t/h	LR I
Pyrolysis	Germany	900°C	Boiler Coupled	10 t/h	LR III
Pyrolysis	Yugoslavia	530°C	Coke Blend	2 units - 1600 t/d	commercial
Cracker	Japan	600°C	Hydrocarbon Cracker	100 t/h (if used for coal)	commercial
Pyrolysis	Germany	750°C	Briquetting	350 t/d	commercial
Pyrolysis	Britain	750°C	Briquetting	700 t/d	commercial

- Low Temperature Carbonization: Formed Coke

The 200 kg/h test facility at Herten, Germany has been used to optimize conditions for liquid yields. The commercial scale plant in Bosnia, Yugoslavia produces a low volatile char and binder from lignite is blended with coking coal to form coke briquettes.

A schematic of the Bosnia plant is shown in Figure 3.14. The raw coal is first conveyed via a chain conveyor (1) to the elevated bunker (2). Four dosing screws (4) operating in parallel feed the coal to four pipes where cooled carbonization gases convey the coal to the reactor (5). Hot coke (530°C) flows from the collecting bin (6) and is mixed with the coal in the reactor. Four blast nozzles and the coke inlet are arranged so as to mix the coal and coke as thoroughly as possible.

Another reactor-mixer developed by LR consists of twin screws rotating in the same direction. This system is used for caking coals and is described in following sections.

Mixing is further enhanced by the release of volatiles from the coal. Gate valves control the flow of coke from the reactor to the lift pipe (7) where it is lifted pneumatically to the collecting bin (6) by air and combustion gases. The lift pipe is heated with carbonization gas (370°C) and by partial combustion of the coke. Supplementary combustion of middle oil brings the lift pipe temperature to 530°C. Surplus char in the collecting bin is continuously withdrawn to the char cooling section. Solid levels are controlled by means of radioactive preparations and Geiger counters.

Waste gas from the collecting bin (6) is cooled to 390°C in the air preheater (8) and then flows through multicyclone (9) and onto a raw coal dryer.

The carbonization gas from the reactor (5) is dedusted at 450°C in a cyclone (10). The gas is cooled to 105°C chiefly by evaporation of flushing tar recycled continuously through the spray cooler (11) and the scrubbing cooler (12). Residual dust is removed in the scrubbing cooler. The tar fog formed in the cooler and scrubber are removed in the electrostatic precipitator (13). The gas is indirectly cooled to 30°C in the cross-tube cooler (14). The middle oil and gas liquor are separated in the separating tank (15). When operating with a lignite there is a surplus of middle oil which is sold. A small percentage of the middle oil is burnt in the lift pipe.

Gas liquor together with waste liquor from the coke oven plant are subject to further treatment. The Phenosolvan method and supplementary biological purification will ensure the most stringent environmental laws are met.

The majority of the cooled gases, containing some vapours, are supplied to either the burner of the lift pipe or to the coke works system. A small portion is supplied to the dosing screws and then recycled to the reactor as injector gas.

Smaller amounts of coke collected in the multicyclone (9) and cyclone (10) are added to the main stream from the collecting bin (6). The coke is cooled via direct contact with water to 100°C. It is then tarred and moistened in a double screw mixer. The total tar product from the precipitation (13) is used along with a portion of the gas liquor for moistening. The semi-coke containing 6% water is then mixed with the coking coal.

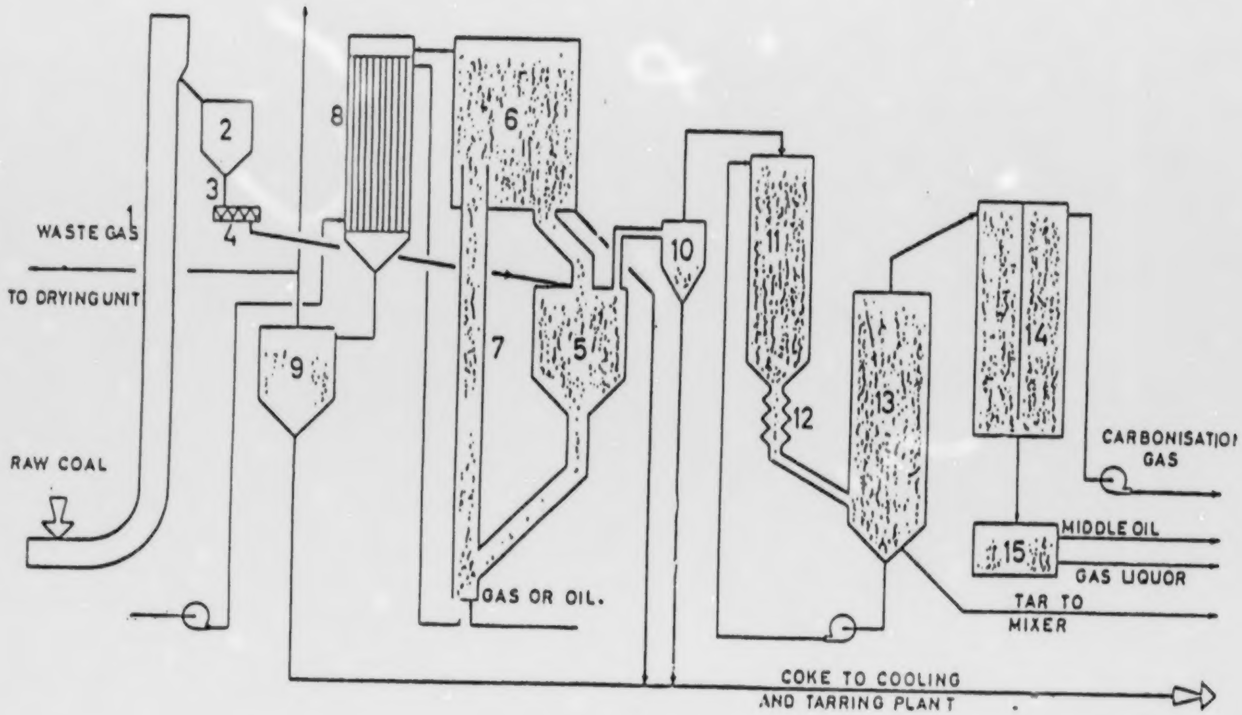

WASTE GAS
1
2
3
4
TO DRYING UNIT
RAW COAL
8
9
6
7
5
10
11
12
13
14
15
GAS OR OIL.
CARBONISATIO
GAS
MIDDLE OIL
GAS LIQUOR
TAR TO
MIXER
COKE TO COOLING
AND TARRING PLANT

Figure 3.14

The Lurgi-Ruhrgas Carbonization Process for Formed Coke Production at Bosnia, Yugoslavia

(Rammler, 1966)

The semi-coke produced serves to reduce the caking properties of the coking coal. The unit is operated to provide maximum amount of semi-coke to the coke ovens thus reducing the amount of imported coking coal used.

When the coal throughput is less than 2/3 of the rated capacity of the plant, steam is added to the air in the lift pipe. This prevents the pneumatically conveyed circulating char from clogging.

A number of factors influence the grain size of the circulated coke. Petrographic composition and increased water content of the coal and increased char circulation rate may increase the fines content of the char. This tends to upset stable pneumatic coke circulation and increase the dust content to the tar. The Bosnia plant operates with 17 to 25% dust content in the tar at 95°C.

Depending upon the devolatilization characteristics of the coal being processed, maximum volatile content of the char and maximum lift tube temperatures must be observed in order to avoid tar release in the lift pipe.

- Low Temperature Carbonization: Char and Tar

The above discussion has dealt with the operation of the Bosnia, Yugoslavia char-coke plant. Further low temperature operational experience has been gained at the Herten, LR I test facility operating at dried coal feeds up to 500 kg/h. This plant has been operated to optimize liquid yields. Table 3.22 shows operational and yield data for the HvAb coal.

The LR I pilot plant has basically the same reactor system as the Bosnia plant with one major difference - the devolatilization of the coal takes place in a twin screw mixer. The clearance between the screws and their housing is small, therefore reducing buildup of caked coal when processing high FSI coals. A schematic of the LR I plant is shown in Figure 3.15. The combustion gases from the lift pipe are dedusted in the cyclone (7) and then routed to an after treatment section (8). The volatiles are removed from the end of the mixing chamber (1) and sent to the gas cooling and condensation train (8).

The condensation train is operated to obtain two and maybe three tar fractions. The aim being to avoid oil/water emulsions which can quickly stop operations. The first stage cools the gas using injection and evaporation of gas liquor and scrubs the gas of dust using a recirculated heavy tar. The remaining tar fractions are virtually dust free. The final stage cools the gas to 30 - 40°C collecting both tar oils and water. These are separated by gravity separation. Gas naphtha is not collected in this system.

Waste gas after-treatment is dependent upon the requirements of each individual case. The flue gas from the circulating system along with entrained dust is cooled to 100 - 200°C. Surplus char is also cooled to about 200°C in this stage. Further char cooling can be carried out in fluidized bed heat exchangers and finally by spraying with a water/gas liquor mixture.

Table 3.22

Material Balance for the LR I Pyrolysis Plant at Herten, West Germany

(Rammler, 1982)

Run No.	1	
Coal	High Volatile	
Type	Bituminous A Coal	
Place of Origin	Germany	
Moisture (%)	3.3	
Ash (%)	5.8	
Volatile Matter (%)	34.6	
Volatile Matter maf (%)	38.1	
Tar (Fischer Assay) (%)	14.7	
Caking Index (Damm)	~ 20	
Grain Size Distribution		
> 2 mm (wt. %)	9.9	
2 - 1 mm (wt. %)	20.0	
1 - 0.5 mm (wt. %)	19.1	
0.5 - 0.2 mm (wt. %)	22.5	
0.2 - 0.1 mm (wt. %)	12.1	
< 0.1 mm (wt. %)	16.4	
Carbonization Temperature (°C)	583	
Products kg/t of Feed Coal		
Circulated Material	511	
Dust in Carbonization Gas	31	
Dust in Flue Gas	104	
Carbon Burnt in Liftpipe	10	656
Coke		
Tar (dust-free)	178	
Oil (dry, dust-free)	19	
Oil in Gas Liquor	2	
Gas Naphtha	7	
Liquid Hydrocarbons		206
% of Fischer Assay	140	
Gas Liquor	81	
Water in Oil	0	
Water in Carbonization Gas	1	
Water		82
Carbonization Gas	51	
Post Devolatilization Gas	5	
Gas		56
Total		1000

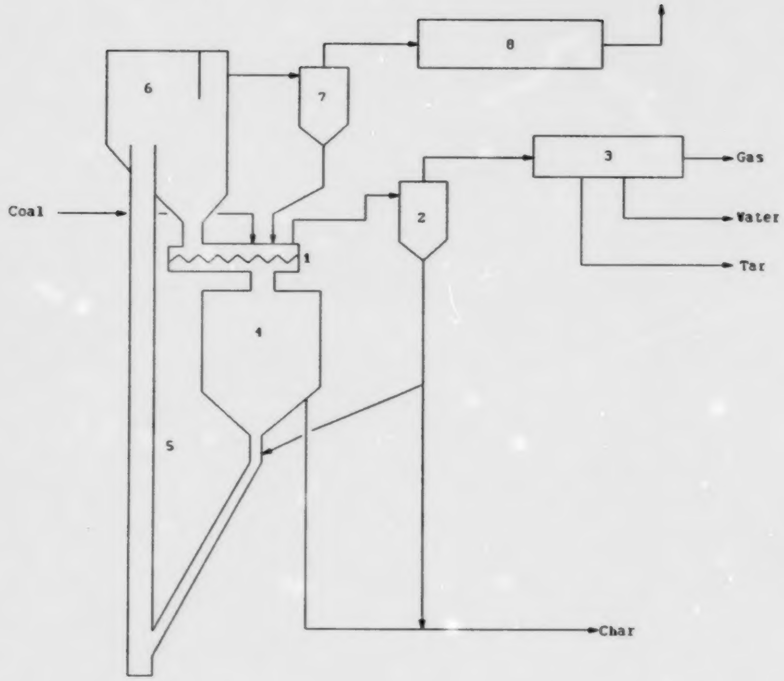

8
6
7
3
Gas
Coal
2
Water
1
Tar
4
5
Char

Figure 3.15

The Lurgi-Ruhrgas LR I Carbonization Process for Formed Coke Production

(Rammler, 1966)

The following recommendations have resulted from the LR I testing:

1. Volatilization should be limited to a few large plants with first stage hydrotreating of the tar to produce transportable fuel. The fuel should be transported to a central plant for final upgrading.

2. The coal should be dried prior to pyrolysis to reduce gas handling. It is much easier to remove water during coal drying than from the gas liquor. Suspension driers with throughputs of 20 to 30 t/h evaporated H_2O are feasible with present day technology.

3. One mixing screw per rise pipe will optimize the liquid yield and reduce the plant height required. A unit capable of processing 250 t/h dried coal would be 80 m high.

4. A number of smaller units (< 250 t/h) would offer reduced operational risk and shutdown losses. Each smaller unit should have its own first condensation stage and feed to a common downstream liquid process unit.

· Yield Data: Char and Tar Operation

Table 3.23 shows the composition of the tars produced from the LR I pilot plant when processing two different sub-bituminous A coals. The presence of calcium during the flash pyrolysis would tend to decrease the heavy tar yields and increase the middle oil and gas naphtha yields. Table 3.24 shows the composition of the carbonization gases from the LR I pilot plant when processing HvAb coal on a dry, N_2 and O_2 free basis. The high composition of CO, H_2, CH_4 and higher molecular weight hydrocarbons make the gas a good source of H_2 for tar hydrotreating (see Table 3.24 for yields).

The total char analysis from the above test is shown in Table 3.25. The ash content has risen from 5.8% in the parent coal to 9.8% in the char.

Comparing grain sizes prior to pyrolysis (Table 3.22) and after pyrolysis (Table 3.25) it can be seen that both the larger (> 1 mm) and smaller (< 0.2 mm) particle size fractions have been reduced and that the mid-size fractions have increased. This will simplify downstream processing such as briquetting or combustion.

· Lurgi-Ruhrgas Electrical Generation

Extensive testing of the LR system operating connected directly to a boiler furnace was done between 1957 and 1960 (Sommers 1974). The final configuration of the LR III plant at Dorsten, Germany is shown in Figure 3.16. This plant processed 10 t/h of dried coal operating typically at 920°C. The very fine grained portion of the char from the collector bin, the dust separated from the product gas and the waste gas from the collector bin were blown directly into the boiler firing chamber through a refractory insulated line.

Maximum operating temperatures in the lift tube are limited by the ash fusion temperature.

Table 3.23

Composition of Low Temperature Tar From the LR I Process

(Rammler, 1982)

Type of Coal Volatile Matter (wt %) Tar Content (wt %)	Sub-Bituminous A Coal 44.6 13.1			Sub-Bituminous A Coal 45.5 14.8		
	Heavy Tar	Middle Oil	Gas Naphtha	Heavy Tar	Middle Oil	Gas Naphtha
Percentage on Total Tar (wt %)	70.7*	22.4	6.9	72.7*	20	7.3
Ultimate Analysis						
Ash (wt %)	1.57	0.16		3.28	0.007	
C (wt %)	81.81	83.36		79.9	83.09	
H (wt %)	5.98	9.76		5.8	8.92	
S (wt %)	0.37	0.43		0.48	0.51	
N (wt %)	1.01	0.33		1.20	0.66	
O (wt %)	9.26	5.96		9.34	6.75	
Density at 20°C (kg/liter)	1.259	0.95	0.738	1.234	1.00	0.728
Dust Content (wt %)	9.3	0.02		6.7	3.83	
Conradson Residue (wt %) (in dust-free tar)	45.3	1.06		45.7	8.91	
Flash Point (°C)	305	61		251	70	
Softening Point (°C) (Krämer-Sarnow)	129			116		
Pour Point (°C)		-19			+ 2	
Viscosity (°E/°C)	6.85/200	1.62/20		6.44/220	5.43/20	
Initial Boiling Point (°C)		85	33		118	
Final Boiling Point (°C)		318	139		329	
Distillate (Vol. %)		89.5	97.0		81	
Last Runnings (Vol. %)		1.5	0.5		1.5	
Residue and Loss (Vol. %)		9.0	2.5		17.5	

* Based on dust-free tar

Table 3.24

Composition of Carbonization Gases for the LR Pyrolysis Of a HvAb Coal (Dry, N_2 and O_2 Free Calculated)

(Rammler, 1982)

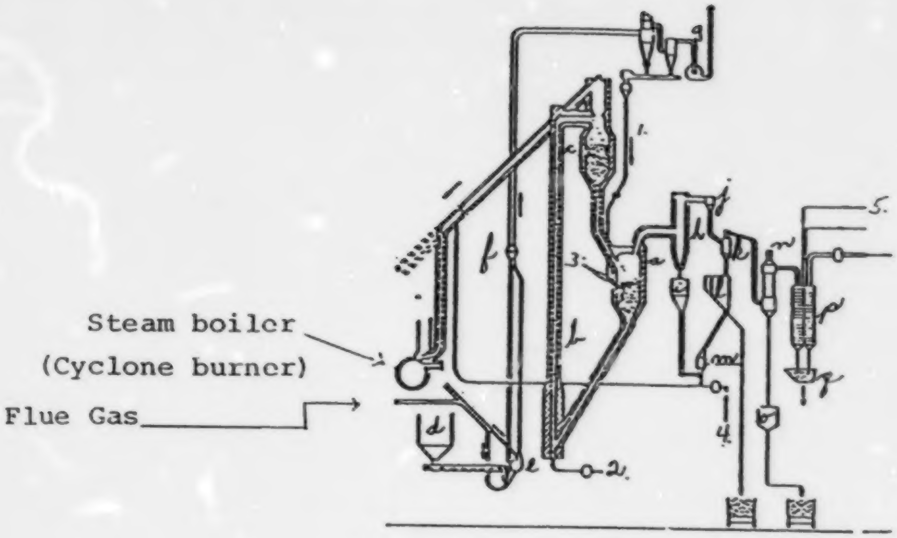
Carbonization Temperature (°C)	583
H_2S (Vol. %)	4.5
CO_2 (Vol. %)	3.9
CO (Vol. %)	11.3
H_2 (Vol. %)	13.2
CH_4 (Vol. %)	49.6
C_2H_4 (Vol. %)	3.1
C_2H_6 (Vol. %)	9.1
C_3H_6 (Vol. %)	2.9
C_3H_8 (Vol. %)	2.4
Σ (Vol. %)	100.0
Density (kg/liter)	0.921
Net Calorific Value (MJ/Nm^3)	34.22
Naphtha (g/Nm^3)	93.9
H_2S (mg/Nm^3)	6920
SO_2 (mg/Nm^3)	-
NH_3 (mg/Nm^3)	1340

Table 3.25

Analytical Data of Total Char for the LR Pyrolysis of a HvAb Coal

(Rammler, 1982)

Carbonization Temperature (°C)	583
Char	
Proximate Analysis	
Ash (wt %)	9.8
Volatile Matter (wt %)	7.4
Fixed Carbon (wt %)	82.8
Ultimate Analysis	
Ash (wt %)	9.8
C (wt %)	81.2
H (wt %)	1.7
S (wt %)	1.1
N (wt %)	1.6
O (wt %)	4.6
Grain-Size Distribution	
> 2 mm	2.2
2.0 - 1.0 mm	6.3
1.0 - 0.5 mm	49.3
0.5 - 0.2 mm	26.4
0.2 - 0.1 mm	6.1
0.1 - 0.063 mm < 0.063 mm	9.7


Steam boiler
(Cyclone burner)
Flue Gas
1.
c
i
f
b
3.
a
j
h
k
5.
n
m
p
q
4
o
d
e
2.

Apparatus:

a. producer shaft
b. rise pipe
c. collector
d. coal bunker
e. coal mill
f. screen
g. coal separator
h. cyclone
i. coke collector

j. injector nozzle
k. washer cyclone
l. washer unit
m. mud drier
n. electro-filter
o. tar collector
p. transverse tube cooler
q. immersion

Inputs:

1. coal(dry)
2. conveyor air
3. steam
4. nitrogen
5. cooling water

Figure 3.16

The Lurgi-Ruhrgas LR III Devolatilization Plant for Coal

(Döreng et al., 1975)

The control of particle sizes fed to the furnace was a problem that was not solved at the completion of testing. The unit originally had a mixer reactor as in the Herten plant but later used a simple jet-coal feeder in the reactor. This feeder resulted in coarse-grained structure and non-uniform quantity of dust leaving the reactor. This was due to its performance depending upon the filling level of the reactor. Intermediate storage of the dust elevated the fluctuating problem somewhat but metering of the dust to the refractory-lined pipe going to the boiler operated irregularly. This problem was not solved.

Particle disintegration takes place primarily in the lift pipe. Careful control of the coal particle size as fed and the lift pipe operating condition were necessary to control the char particle size.

Based on the LR III test results the predicted heat and material throughputs for a commercial operation are shown in Table 3.26.

The pyrolysis boiler system operated smoothly but optimum solutions to the above problem were not found.

Protopapas et al. (1987) has prepared a summary of char producing processes for the USDOE.

Of the twelve externally heated fixed bed type reactors reviewed, only one facility was in operation (for the Protopapas report, all facilities referred to will be pilot to commercial scale). That facility is the Coalite Chemical Co. in England. The plant processed approximately 4000 t/d of coal to produce 75% smokeless fuel for the domestic market.

A total of seventeen vertical moving bed processes were reviewed. The only active facility was the Stanford Research Institute unit at Gillette, WY, in which approximately 4000 t/d of sub-bituminous coal is processed to 50% boiler fuel.

In the horizontal externally heated moving bed category, ten processes were reported on, all were inactive.

The Synfuels Genesis, Inc. 100 t/d rotary tray moving bed type reactor system was reported under construction. The plant produced 75% boiler fuel. Two additional reactors of similar design were abandoned.

Nine fluidized bed type reactors were all reported as inactive. The largest in this category is the COED process that FMC holds the licence rights to.

Five entrained bed reactors were reported on. Only the Lurgi-Ruhrgas process in Scunthorpe, UK was active, processing 700 t/d to produce 64% boiler fuel and 10% chemicals.

No commercial scale hydropyrolysis processes were reported on.

Since the release of Protopapas's report, the USDOE has begun to fund a number of partial gasification projects in the US that are discussed later in this report.

Table 3.26

Heat and Material Throughputs - LR Devolatilizing of Gas Flame Coal (22% Ash, 8% H_2O) in Joint Operation with Boiler

(Döring et al., 1975)

Quantity and Heat Balance	990/920°C	
Products	Weight	Calorific Value
Char to Boiler, % of Raw Coal	54.0	48.5
Burned Char, % of Raw Coal	6.9	11.5
Dry Mud, % of Raw Coal	6.1	6.2
Pitch and Tar, % of Raw Coal	3.8	6.1
Medium Oil and Naphthalene, % of Raw Coal	0.8	1.6
Light Oil, % of Raw Coal	1.5	2.6
Gas without Petrol and H_2S, % of Raw Coal	15.7	23.0
$H_2S + NH_3 + H_2O$, % of Raw Coal	11.2	0.5
Residual Element (Coking Products), % of Raw Coal	--	--
Supply to Boiler	at 900°C	
Flue Gas, % of Raw Coal	77.2	2.0 + 3.6
Water Vapour, % of Raw Coal	1.8	0.2
Char, % of Raw Coal	54.0	48.5 + 2.4
TOTAL	133.0	56.7
Heat of Flue Gas From Boiler, % of Raw Coal	92.5	2.9
In Coal Upon Drying, % of Raw Coal	93.5	0.4
Heat Lost in Water Vapour, % of Raw Coal	99	1.2
For Pre-Heat (from 60 to 450°C), % of Raw Coal	72.7	1.3
Utilization of Coal (Summary)		
Burner Gas and Hydrocarbons, % of Raw Coal	33.3	
Steam Boiler (Supply Less Requirements), % of Raw Coal	60.0	
Waste Heat Stream/Losses, % of Raw Coal	0/6.7	

3.3.2 Active Processes

This section of the report deals with the partial gasification processes that are currently active. The data are presented in two sections, one dealing with those processes that produce electricity and the second examines those that produce co-products such as gas, liquids and char.

3.3.2.1 Review of Active Electrical Generation Partial Gasification Processes

The application of partial gasification of coal to the generation of electricity has been pursued due to the potential of achieving higher cycle efficiencies. The cycle efficiency can be defined as the percentage of the chemical energy stored in the molecular bonds in the coal that is converted to electrical energy (which is available to the grid).

Partial gasification of coal for electrical generation would utilize the following general steps:

1. *Gasifier*: Coal is partially gasified resulting in the production of solids containing mostly carbon and any inert material such as ash, and combustible molecular species that are in the gas phase under the conditions of operation. Some energy recovery may take place here in the form of the generation of steam to be used in a *steam turbine*.

2. *Combustor*: The solid material is sent to a combustor, typically a circulating fluidized bed at atmospheric or elevated pressure, where the remainder of the chemical energy is released. Steam is raised for use in a *steam turbine*.

3. *Gas Turbine*: The gas phase combustible material from the gasifier is combusted and expanded through a gas turbine. The turbine exhaust is sent to the combustor.

4. *Steam Turbine*: Steam generated in the gasifier heat recovery system, the combustor and possibly the gas turbine are utilized.

One of the advantages of the partial gasification process over the total gasification process is that the gasifier reactor costs are lower. The initial volatilization and gasification steps of coal are achieved relatively easily. As the gasification process progresses, the remaining char becomes more refractory and the gasification becomes more difficult. The conversion of the remaining carbon can be carried out much easier under the oxidizing combustion conditions. An optimum can be achieved between the gasifier and combustor sizing.

The optimization of the operation and sizing of the steam and gas turbines can also be carried out in conjunction with the combustor and gasifier sizing.

The optimum sizing of the gasifier, combustor and turbines results in the prediction of cycle efficiencies in the 45% range. This compares to efficiencies in the 41% range for total gasification. Total gasification technologies will achieve higher efficiencies with the development and application of hot gas clean up (sulphur, particulates, etc.)

The application of partial gasification to the generation of electricity is presently being pursued by a number of companies. Listed below are the companies whose processes have been discussed in this report.

Table 3.27

Companies Involved in the Evaluation of Partial Gasification of Coal As Applied to Electricity Generation

Process Developer	Reactor Type	Scale (kg/h)
British Coal	Spouted Bed	500
M. W. Kellogg	Entrained	42
VEW	Entrained	10,000

British Coal

· Process Description

British Coal has started a project called the Grimethorpe Topping Cycle Project (GTCP) (British Coal Corporation (1990)). The objective of the GTCP is to investigate the feasibility of operating a commercial gas turbine on coal-derived flue gas at elevated temperatures. There are a number of facets of the project including:

- generation of flue gas from a pressurized fluidized bed (PFB)
- testing a turbine in the cleaned gas stream from the PFB after additional heating by propane firing
- testing additional turbine blade material in a separate side stream and high temperature/pressure filter performance testing

Associated research will be carried out at the Cheltenham facility on pressurized spouted fluid bed gasification.

The British Coal, Coal Research Establishment at Cheltenham has developed a spouted fluid bed gasifier designed to produce low heating value (3.5 to 4 MJ/m^3, gross) gas. The development spanned the period 1974 to 1989. The process converts approximately 60 to 70% of the coal and was developed to supply industrial markets (Gale et al., 1989). The desire for higher carbon to gas conversions (90%) has led to the development of a gasifier using a bed of inert material and reduced coal to air ratios. The use of the inert bed material allows a stable bed depth to be maintained. The lower conversion variant mode of operation (in a pressurized version) would be applied to electrical generation.

Figures 3.17 and 3.18 show the two proposed variations of the British Coal "Topping Cycle" for electricity generation. In both variations, a pressurized spouted fluidized (PSFB) bed gasifier is used. Char from the base of the gasifier and particulate material removed from the gas stream are fed to a combustor. One variation uses a circulating fluidized bed combustor (CFBC) and the other a pressurized bubbling fluidized bed (PBFBC) for char combustion. The projected cycle efficiency for the CFBC configuration is 44.8% (% HHV) and 44.3% for the PFBC (Dawes et al., undated). The CFBC option is expected to have a slightly higher capital cost/kWe. Both variations utilize a combination of gas and steam turbines to generate electricity.

Discussion of the Topping Cycle operation will be based on experience with the atmospheric gasifier discussed above. Pressurized gasification tests are just beginning at the time of this writing.

Gale et al. (1989) describes the development of the spouted bed gasifier. Coal will be fed to the gasifier via a series of lock hoppers. The bed utilizes a coal derived char bed while converting approximately 60 to 70%, mass basis, of the coal to gas. The reactor is shown in Figure 3.19. The majority of the fluidizing gas is introduced as a jet at the apex of the conical base. Gas velocities in the jet are in the range of 25 m/s. Additional fluidizing gas can be injected in the conical section to ensure complete fluidization. Solids circulation caused because of the gas flow and high velocities in the jet prevent agglomeration of caking coal particles.

Dawes et al. gives the following reason for the choice of a fluidized bed reactor:

- operation at 1000°C results in a limited amount of tar being produced, avoids the high alkali contents that are seen at the higher temperatures used in entrained reactors and is below the temperature where the ash becomes sticky
- bed temperature, particle residence time and good gas solids contacting allow sulphur retention to be carried out in the reactor
- the above two factors allow operation at higher efficiencies due to the elimination of lower temperature gas cleaning systems
- the fluidized bed reactor is tolerant to variations in the coal particle size and composition

Bed heights of 2 m are used, which is above the maximum spouted bed height, resulting in the upper section of the bed behaving in a slugging mode. Two coal feed sites may be used, either at the base of the cone or via a screw feeder above the bed.

An air blown gasifier was chosen in order to reduce costs and because it can produce gas qualities that are acceptable for gas turbine utilization.

Work is currently under way to develop a pressurized version of the British Coal fluidized gasifier. Figure 3.20 shows an isometric view of the pilot plant that is under construction. The gasifier will have a spouted bed section that is 0.34 m ID and 6 m high. The freeboard section, on top of the reactor, will be 0.45 m ID and 10 m high. The reactor will operate at 1000°C and 1 to 1.5 MPa. The goal is to produce 3.6 MJ/m^3 (HHV) gas. Coal conversions of 60 to 80% are considered optimum for the Topping Cycle operation.

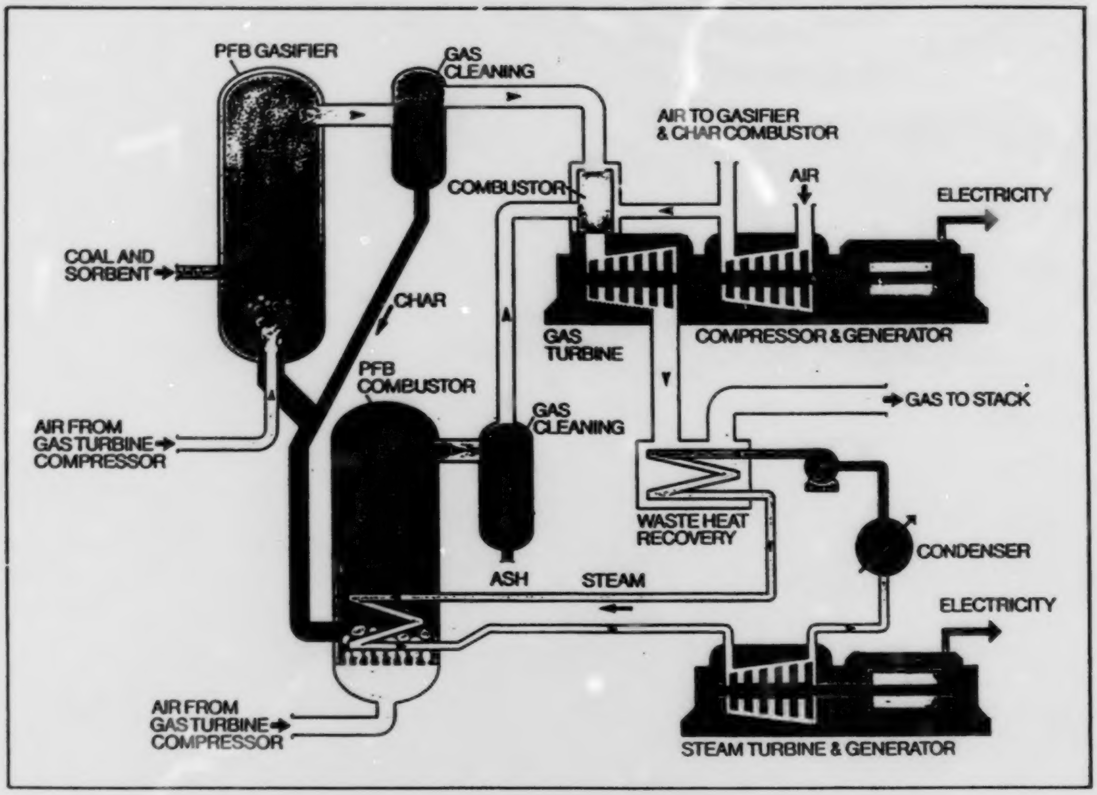

PFB GASIFIER
GAS CLEANING
AIR TO GASIFIER & CHAR COMBUSTOR
COMBUSTOR
AIR
ELECTRICITY
COAL AND SORBENT
CHAR
GAS TURBINE
COMPRESSOR & GENERATOR
PFB COMBUSTOR
GAS TO STACK
AIR FROM GAS TURBINE COMPRESSOR
GAS CLEANING
WASTE HEAT RECOVERY
CONDENSER
ASH
STEAM
ELECTRICITY
AIR FROM GAS TURBINE COMPRESSOR
STEAM TURBINE & GENERATOR

Figure 3.17

The British Coal Topping Cycle Based on a Pressurized Fluidized Bed Combustor

(Dawes)

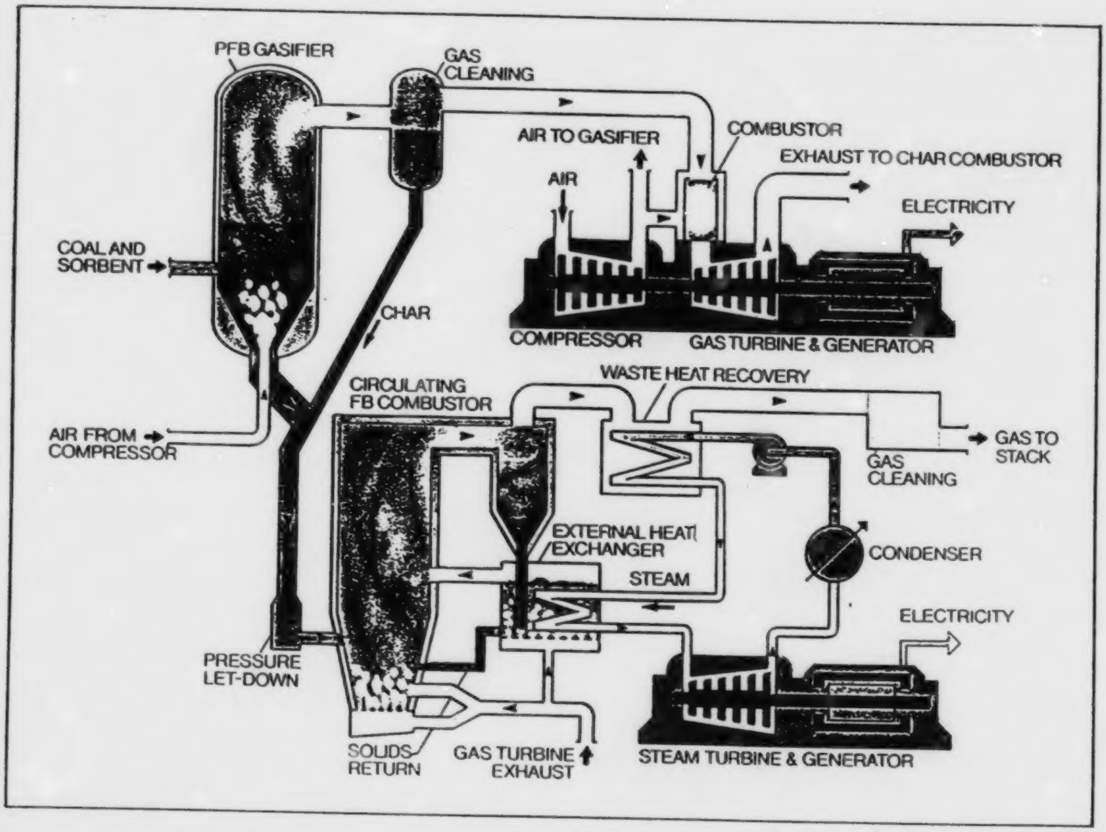

PFB GASIFIER
GAS CLEANING
AIR TO GASIFIER
COMBUSTOR
EXHAUST TO CHAR COMBUSTOR
AIR
ELECTRICITY
COAL AND SORBENT
CHAR
COMPRESSOR
GAS TURBINE & GENERATOR
WASTE HEAT RECOVERY
CIRCULATING FB COMBUSTOR
AIR FROM COMPRESSOR
GAS TO STACK
GAS CLEANING
EXTERNAL HEAT EXCHANGER
CONDENSER
STEAM
ELECTRICITY
PRESSURE LET-DOWN
SOLIDS RETURN
GAS TURBINE EXHAUST
STEAM TURBINE & GENERATOR

Figure 3.18

The British Coal Topping Cycle Based on a Circulating Fluidized Bed Combustor

(Dawes)

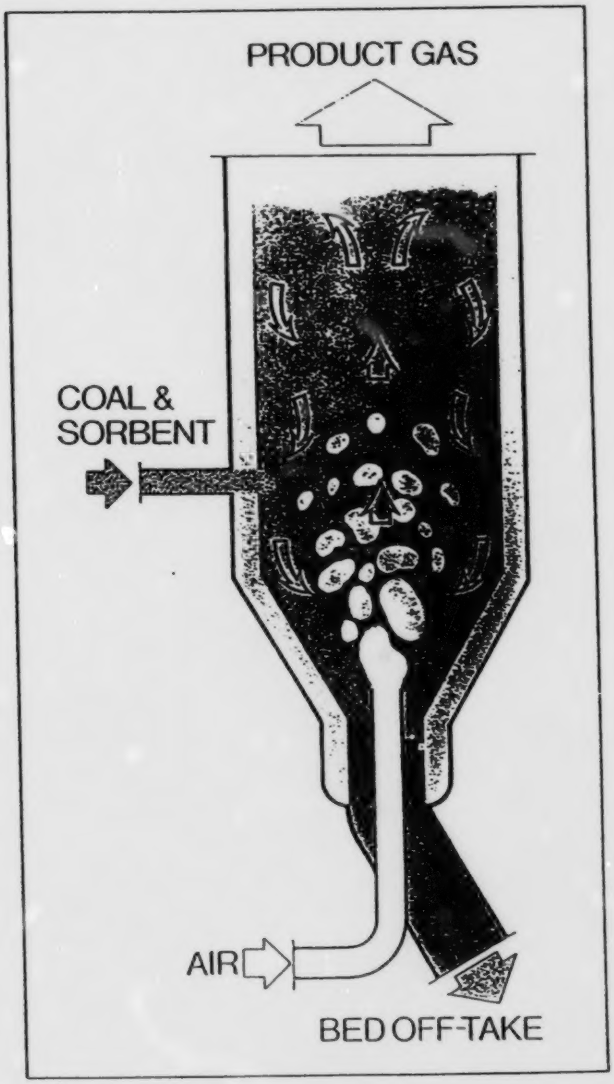

PRODUCT GAS
COAL & SORBENT
AIR
BED OFF-TAKE

Figure 3.19

The British Coal Spouted Bed Gasifier Reactor

(Dawes)

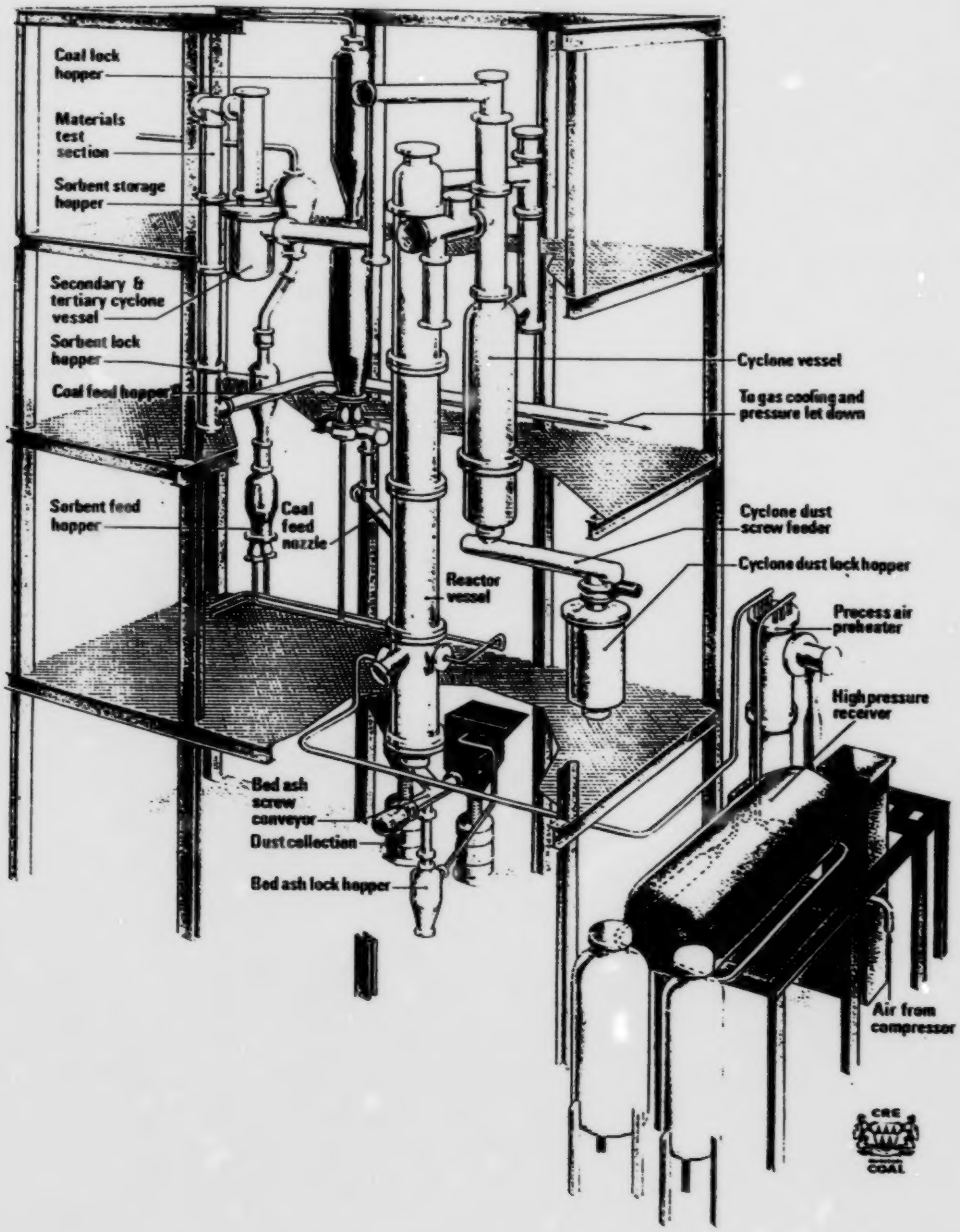

Coal lock hopper
Materials test section
Sorbent storage hopper
Secondary & tertiary cyclone vessel
Sorbent lock hopper
Coal feed hopper
Sorbent feed hopper
Coal feed nozzle
Reactor vessel
Bed ash screw conveyor
Dust collection
Bed ash lock hopper
Cyclone vessel
To gas cooling and pressure let down
Cyclone dust screw feeder
Cyclone dust lock hopper
Process air preheater
High pressure receiver
Air from compressor
CRE
COAL

Figure 3.20

An Isometric View of the British Coal 0.5 tonne/h Pressurized Fluidized Bed Gasification Pilot Plant

(Dawes)

The CaS produced under the reducing environment in the gasifier must be oxidized to $CaSO_4$ prior to disposal.

Successful operation of the 500 kg/h gasifier at Cheltenham will result in the building of a 10 t/h pressurized gasifier at Grimethorpe (Dawes et al., 1989). This facility will see the integration of gasifier and combustor. British Coal plans to have a demonstration plant in operation by 1995 and a full scale facility designed by 2000.

The majority of the hot gas cleanup studies are done at British Coal's Grimethorpe (British Coal/CEGB, 1988) facility. Particulate loadings of < 10 ppm is the target level. Ceramic filters are expected to remove particulates in micron range and up. Submicron particles tend to follow gas stream lines in turbines thereby mitigating potential erosion. At the Cheltenham facility, where the pressurized gasifier is being readied, ceramic candle filters are being installed also. These filters will clean gas from the pressurized gasifier.

Corrosion causing species such as sodium, potassium, calcium, chloride and SO_x are of particular concern to turbine blade materials. A 950°C clay sorbent system has been developed in conjunction with the University of Surrey at a laboratory scale. Larger scale tests are planned.

British Coal feels that turbines designed for use on coal-derived gases need further development. Various blade materials and coatings will be tested at the Grimethorpe facility. A single stage Ruston TB 5000 gas turbine will be used. Gas from a pressurized CFB will have its temperature raised with propane firing to simulate the coal generated gas from the pressurized FB gasifier. The CFB gas will have hot gas cleanup carried out on it prior to entering the turbine.

The objective is to operate the gas turbines at 1260°C and higher. Initial testing may be carried out at 1100°C which is below the ash fusion temperature. Investigations into blade cooling are continuing which would allow operation at even higher gas temperatures (i.e. 1400 to 1500°C). Ceramic materials are showing increasing promise. It is estimated that the commercial version of the high temperature turbines may not be available until after the year 2000.

British Coal, Coal Research Establishment at Cheltenham has a number of facilities available for research test work:

- thermogravimetric balance, 1100°C, 10 MPa
- a 5 kg/h throughput, 0.15 m diameter atmospheric pressure gasification test facility for short duration tests,
- a 25 to 60 kg/h throughput, 0.3 m in diameter atmospheric pressure unit to carry out extended tests,
- a 500 kg/h pilot plant used for commercial trials
- a second 500 kg/h unit is undergoing modification to allow operation at 2 MPa and 1000°C.

Significant efforts have been put into the development of computer models of the PFBC and PSFB gasifier operations.

British Coal has extensive experience in the area of pressurized fluid bed combustion of coal (British Coal/CEGB (1988)). Figure 3.21 shows a schematic of the Grimethorpe PFBC facility. Areas such as high pressure feeding of coal, high pressure fluid bed reactor control, high pressure/temperature gas cleaning, gas turbine operation with coal derived gas, materials performance evaluation and solid waste utilization will be valuable in the development of the pressurized gasifier and its application to the Topping Cycle.

M. W. Kellogg

The M. W. Kellogg Company has considerable experience in the fields of coal combustion and gasification (Campbell 1988, Banchik et al., 1989). The KRW coal gasification process has been fully developed and is ready for commercial demonstration.

Banchik et al. (1989) notes that gasification and combustion combined to produce electricity is optimized when the coal is lower in quality or cheapest.

- Process Description

M. W. Kellogg's partial gasification process (Transport Gasifier) is in the development stage. Figure 3.22 shows the M. W. Kellogg 42 kg/h, 9 m high atmospheric pressure transport reactor system, which is under construction.

A schematic of the M. W. Kellogg transport combustor is shown in Figure 3.23. M. W. Kellogg's considerable experience in the field of fluid catalytic cracking and heavy oil cracking is applied to the design of the transport reactor that is used for combustion and gasification. Coal and limestone are introduced at the expanded base of the reactor where they contact the hot circulating limestone and ash. Combustion and sulphur removal take place as the heated air carries (transports) the solids up the riser. Solids and gas are separated in the disengager. Solids carried over from the disengager enter two stages of cyclones. The solids collected in the disengager are collected in the base as a fluidized bed. Some heat recovery takes place in the disengager fluid bed section. No heat exchange takes place in the riser reactor section. Erosion is low due to the completely oxidized state of the particles and the low fluidizing velocity (0.15 m/s). Turndowns are reported to be high due to the higher velocities used in the riser reactor.

A 9.1 m high gasifier is under construction. A 2 tonne/h facility is being designed.

A USDOE funded contract has been carried out to develop a base case flow sheet for the generation of electricity utilizing a partial gasification step (Cho et al., 1990). The plant would have nominally 500 MWe capacity utilizing two or three modules. Both the gasifier and char combustor would use pressurized transport type reactors. A wide variety of parametric studies have been carried out to determine the impact of the process variables on the gasifier and combustor performance as well as the overall plant efficiency.

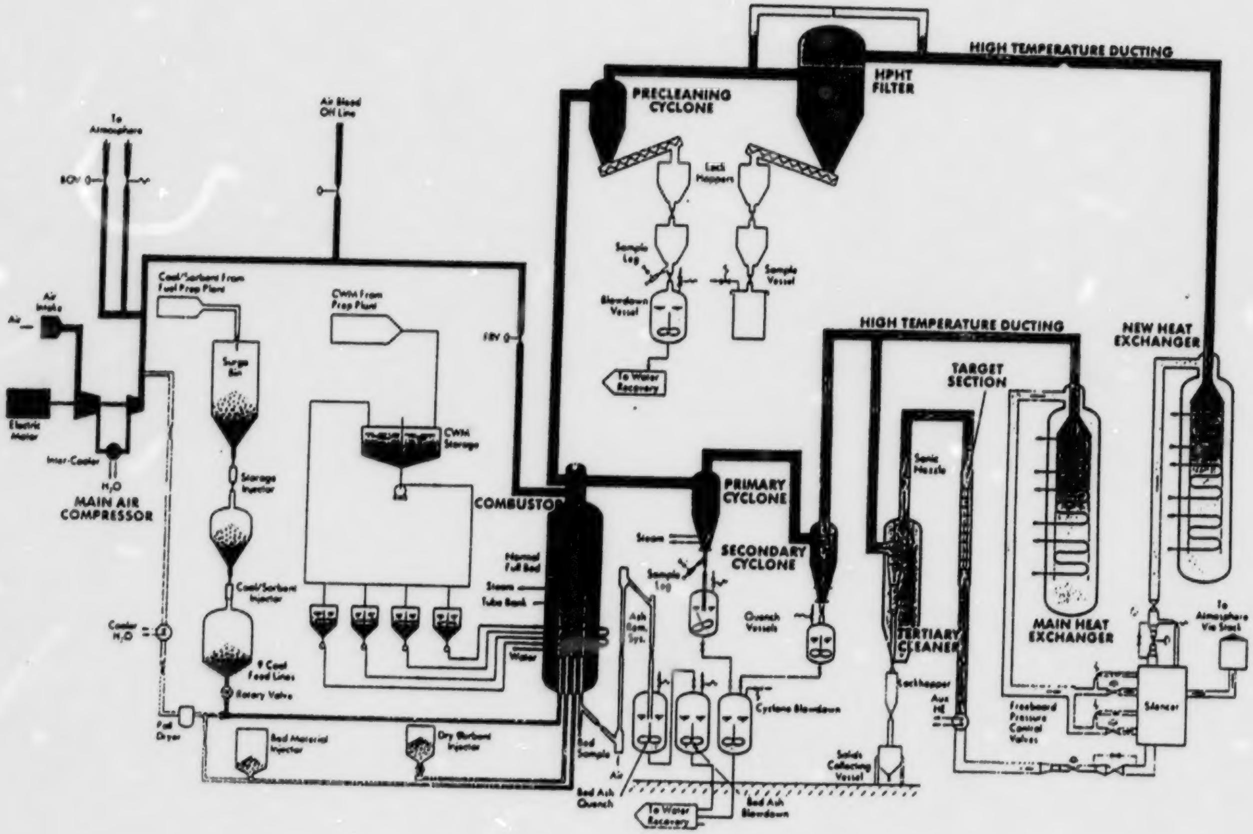

HIGH TEMPERATURE DUCTING
HPHT FILTER
PRECLEANING CYCLONE
Lock Hoppers
Sample Leg
Sample Vessel
Blowdown Vessel
To Water Recovery
To Atmosphere
BOV
Air Bleed Off Line
Air Intake
Air
Electric Motor
Inter-Cooler
H₂O
MAIN AIR COMPRESSOR
Coal/Sorbent From Fuel Prep Plant
CWM From Prep Plant
Surge Bin
CWM Storage
FBV
Storage Injector
Coal/Sorbent Injector
9 Coal Feed Lines
Rotary Valve
Cooler H₂O
Fall Dryer
Bed Material Injector
Dry Sorbent Injector
COMBUSTOR
Normal Full Bed
Steam
Tube Bank
Water
Bed Sample
Air
Bed Ash Quench
Ash Rem. Sys.
PRIMARY CYCLONE
SECONDARY CYCLONE
Steam
Sample Leg
Quench Vessels
Cyclone Blowdown
To Water Recovery
Bed Ash Blowdown
HIGH TEMPERATURE DUCTING
Sonic Nozzle
TERTIARY CLEANER
Lockhopper
Solids Collecting Vessel
TARGET SECTION
Aux HE
MAIN HEAT EXCHANGER
NEW HEAT EXCHANGER
Freeboard Pressure Control Valves
Silencer
To Atmosphere Via Stack

Figure 3.21

The Grimethorpe Pressurized Fluidized Bed Combustion Facility (British Coal)

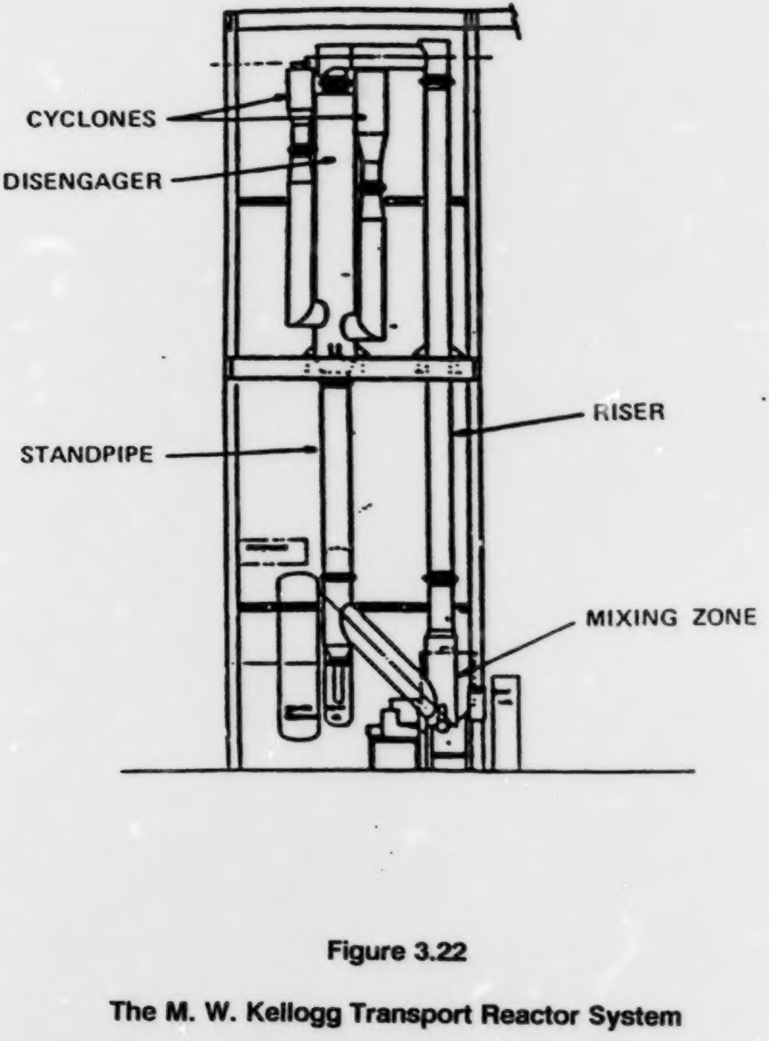
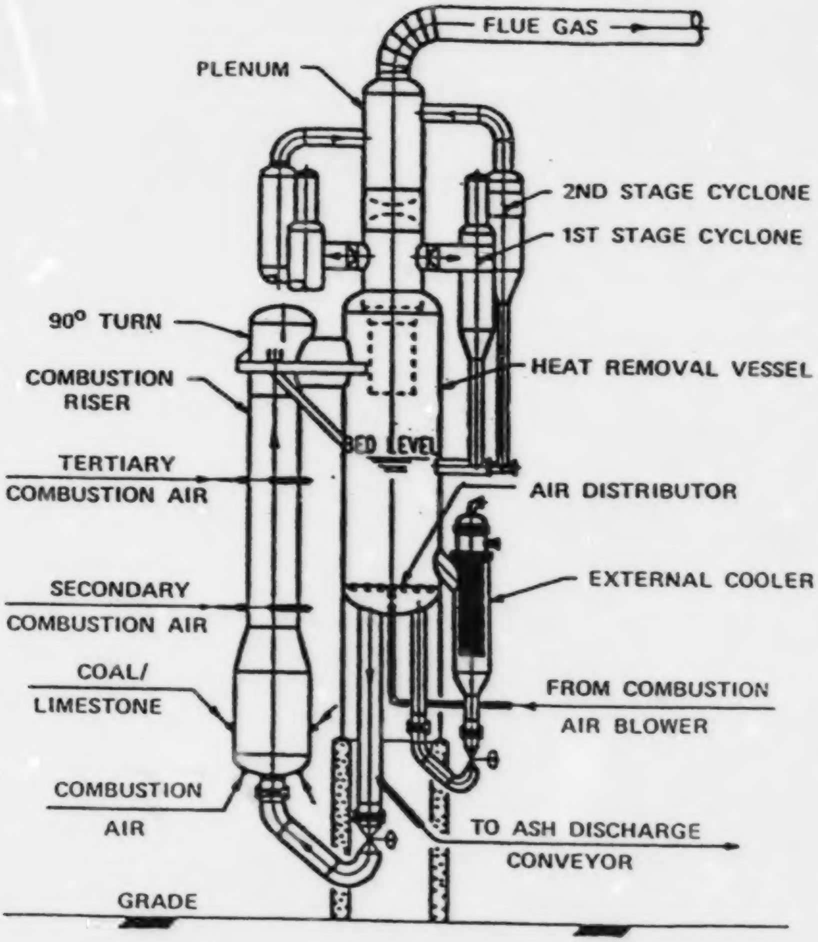

CYCLONES
DISENGAGER
RISER
STANDPIPE
MIXING ZONE

Figure 3.22

The M. W. Kellogg Transport Reactor System

(Banchik et al., 1989)



FLUE GAS
PLENUM
2ND STAGE CYCLONE
1ST STAGE CYCLONE
90° TURN
COMBUSTION RISER
HEAT REMOVAL VESSEL
BED LEVEL
TERTIARY COMBUSTION AIR
AIR DISTRIBUTOR
EXTERNAL COOLER
SECONDARY COMBUSTION AIR
COAL/ LIMESTONE
FROM COMBUSTION AIR BLOWER
COMBUSTION AIR
TO ASH DISCHARGE CONVEYOR
GRADE

Figure 3.23

Design of the M. W. Kellogg Transport Combustor Based Upon a Typical Heavy Oil Cracking Unit Analogy

(Banchik et al., 1989)

The base case design had 551 MWe capacity using two gasifiers/combustors/gas turbines/heat recovery steam generator (HRSG) modules, with a single steam turbine. The gas turbine selected will not be available until 1993 (United Technologies Model V84.4 turbine). The coal used is an Illinois No. 6, HHV 26.6 MJ/kg with 2.51% sulphur as received. A 90% calcium carbonated limestone was selected as a sulphur sorbent.

Figure 3.24 shows a simplified flow diagram of the M. W. Kellogg partial gasification electrical generation process.

Coal is dried to 3% moisture and ground to 50 μm average particle size. The limestone is ground to an average particle size of 100 μm. The coal and limestone are injected into the bed along with steam extracted from the gas turbine and air from a booster compressor and recycled solids. The gasifier operates at 788°C and 1.82 MPa. Approximately 73% of the coal is gasified to produce a gas with a LHV of 6.0 MJ/m^3. The gas goes to a two stage cyclone. After cooling to 400°C, the gas flows through a ceramic barrier filter before entering the gas turbine burner. Most of the collected solids are recycled to the gasifier. The net make of solids (char and partially sulfided sorbent) is sent to the circulating fluidized bed combustor (CFBC) which operates at 890°C and 1.65 MPa. The net make of solids from the CFBC is used to reheat the condensate from the steam cycle and boiler feed water makeup and then discarded.

Solids from the CFBC are removed in a two stage cyclone and a ceramic barrier filter. Heat of combustion is recovered and used to generate steam, superheat and reheat. The solids are recycled to the CFBC. The cleaned combustor gas flows to the gas turbine where it combines with the burned fuel gas and main air supply from the gas turbine compressor. The combined flue gases generate about 296 MWe of net power. Flue gas leaves the turbine at 590°C and enters the heat recovery steam generator, where heat is generated and partially superheated, before being vented to the stack at 138°C.

In the heat recovery steam generator, 16.5 MPa steam is generated and partially superheated to 427°C. Further superheat to 538°C is achieved in the CFBC and the steam sent to the turbine. The intermediate pressure steam from the turbine is reheated to 454°C in the CFBC loop and returned to the steam turbine. Some steam is removed from the turbine at various stages to provide gasifier feed, deaerator steam and condensate preheat. Most of the steam is expanded to 5 cm of Hg with a net generation of 277 MWe from the turbine.

Tables 3.28 and 3.29 show the plant performance and expected emissions.

Capital and operating costs were estimated for the base case following the EPRI (1982) guide. Operating costs were calculated on a ten year levelized basis. Table 3.30 shows the results of the economic analysis. The total plant cost, which includes interest incurred during construction, is US$1017/kW (mid 1990). This is equivalent to a cost of electricity of 56.5 mills/kWh.

Parametric studies were carried out on the gasifier operational conditions (temperature, particle size, gas velocity, pressure, bed density and riser height), combustor (temperature, particle size, air staging and bed density), and the cycle studies (gas turbine temperature, steam cycles, gasifier conversion, air preheat, systems pressure and turndown). The operational conditions reported above are the optimum of the parameters mentioned above.

Table 3.28

M.W. Kellogg Conceptual Partial Gasification Plant Performance Data for the Base Case Study

(Cho et al., 1990)

Plant Input		Plant Output	
Coal to Plant, tph	164	Flue Gas, tph	3,059
Limestone to Plant, tph	22	Waste Solids, tph	38
Steam to Gasifiers, tph	52	Gas Turbine Power, MW	296
Air to Turbine, tph	2,859	Steam Turbine Power, MW	277
Air to Gasifiers, tph	280	Parasitic Losses, MW	22
Air to Combustors, tph	565	Net Power Output, MW	551
Heat Rate (HHV)		7.99 MJ/kWh (45.04% efficiency)	

Table 3.29

M.W. Kellogg Conceptual Partial Gasification Plant Emissions Data for the Base Case Study

(Cho et al., 1990)

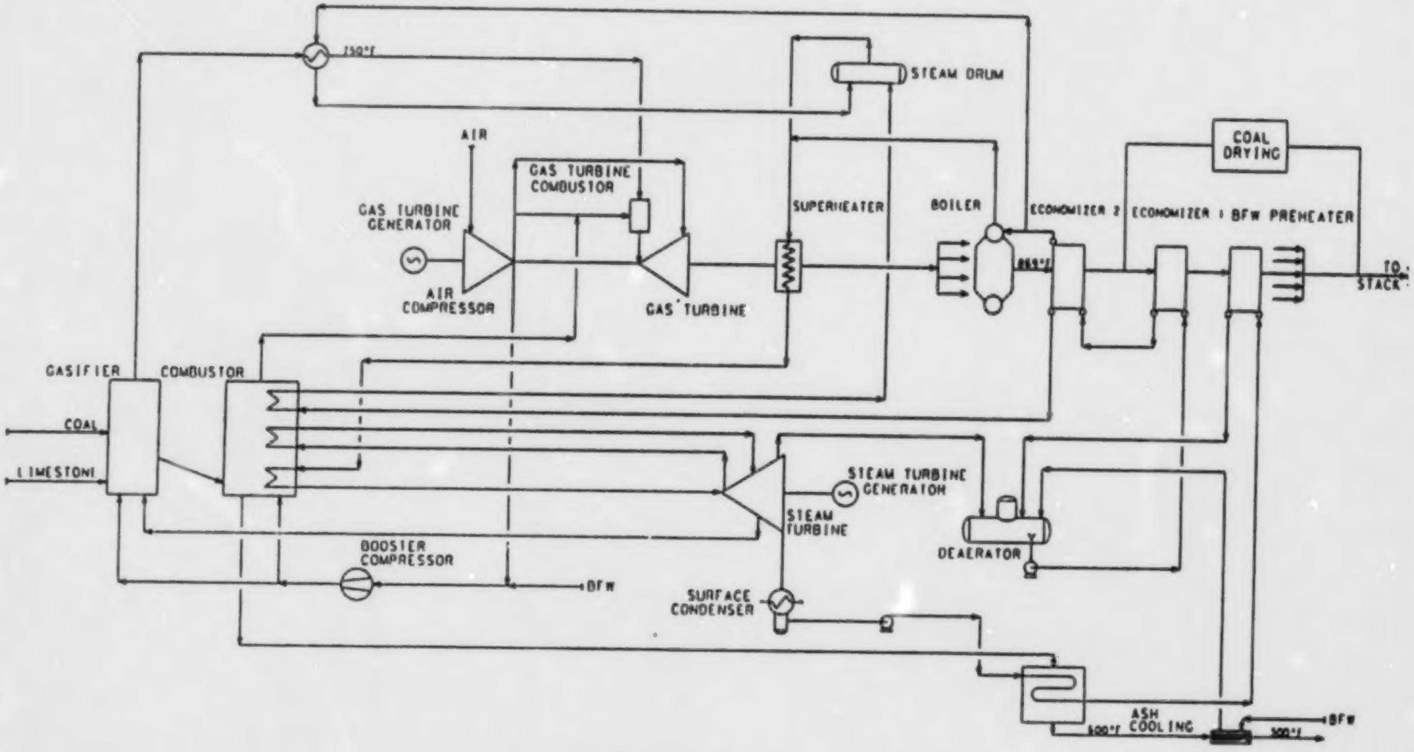
Pollutant	Emissions (g/MJ)	US New Source Performance Standards (g/MJ)
SO_2	0.12	0.52
NO_x	0.03	0.04
CO	0.03	
Particulates	0.006	0.013

Table 3.30

M.W. Kellogg Conceptual Partial Gasification Plant Economics

(Cho et al., 1990)

Investment	US $MM	Cost of Electricity	mills/kWh
Total Plant Cost	491.7	Capital Charges	31.26
Interest Charges During Construction	41.3	Fixed O & M	5.00
Production Costs	14.3	Variable O & M	2.69
Inventory, Spare Parts	13.2	Consumables/Disposal	2.26
		Fuels	15.29
Total Capital Requirement	560.5	COE (10 year levelized)	56.50
	(US $1017/kW) (mid 1990)		



STEAM DRUM
150°F
AIR
GAS TURBINE COMBUSTOR
GAS TURBINE GENERATOR
AIR COMPRESSOR
GAS TURBINE
SUPERHEATER
BOILER
869°F
ECONOMIZER 2
ECONOMIZER 1
BFW PREHEATER
COAL DRYING
TO STACK
GASIFIER
COAL
LIMESTONE
COMBUSTOR
STEAM TURBINE GENERATOR
STEAM TURBINE
DEAERATOR
BOOSTER COMPRESSOR
BFW
SURFACE CONDENSER
ASH COOLING
600°F
300°F
BFW

Figure 3.24

Simplified Process Flow Diagram of the M. W. Kellogg Conceptual Gasification Process

(Cho et al., 1990)

Sensitivity analyses were carried out on variations to the plant design. Variations in the number of trains, dolomite substituted as a sulphur sorbent, lignite feed, 100 µm coal size, 844°C gasifier and slurry feed were studied. All resulted in less than 10% variation in the ten year levelized cost of electricity.

Southern Company Services

The Southern Company Services is an arm of the Southern Company (utility) and has evaluated the application of coal gasification to base load generation of electricity (Gallaspy et al., 1990). A number of gasification schemes were evaluated as installed at a Georgia Power Company's existing installation.

One of the generation schemes utilized a mild gasifier. The gasifier was of the M.W. Kellogg transport pyrolysis design that is discussed in this report. The configuration would have the char from the mild gasifier feed to a 400 MWe air-blown IGCC plant with KRW transport gasification, in-bed desulphurization, two GE MS7001F combustion turbines and a single reheat steam turbine. The liquids from the mild gasification process would be utilized in a peaking load turbine.

A thermal efficiency of 41% (based on the HHV of the coal) was assumed. The total capital required was US$ 521 MM (1230 $/kW of capacity). Net power output was 423.6 MWe. Fixed operating and maintenance costs were 30.1 US$/kw-yr. Variable operating and maintenance costs were 1.05 mils/kWh and fuel costs 12.5 mils/kWh. The ten year levelized cost of electricity in current US$ was calculated to be 61 mils/kWh.

Vereinigte Elektrizitätswerke Westfalen (VEW)

· Process Description

VEW's activity in the field of gasification began with a 1 t/h, 101 kPa, entrained bed reactor facility (Weinzierl 1985) that operated until approximately 1980. Information gained from the operation of the 1 t/h plant was used to design the 10 t/h reactor (Oxley et al. 1987). The 10 t/h, 2 MPa, 1600°C, gasifier began operation in 1984 and has been operational since then. The gasifier and associated facility are integrated into the electrical generation facilities of VEW as shown in Figure 3.25. Milled coal is fed to the gasifier along with preheated air. The cooled char from the gasifier is combusted in a conventional pulverized coal fired design boiler. Figure 3.26 shows a schematic of the VEW gasification facility. More detail of the various steps is given below.

Coal is ground to < 90 µm in a conventional mill. Coal particles are separated from the milling air using a bag filter (Figure 3.26). The coal is brought to the reactor pressure using a nitrogen pressurized system prior to feeding to the metering system. The nitrogen is recovered and returned to the nitrogen compressor. A star feeder is used to send the coal to the air entrainment feed system that carries it to the gasifier.

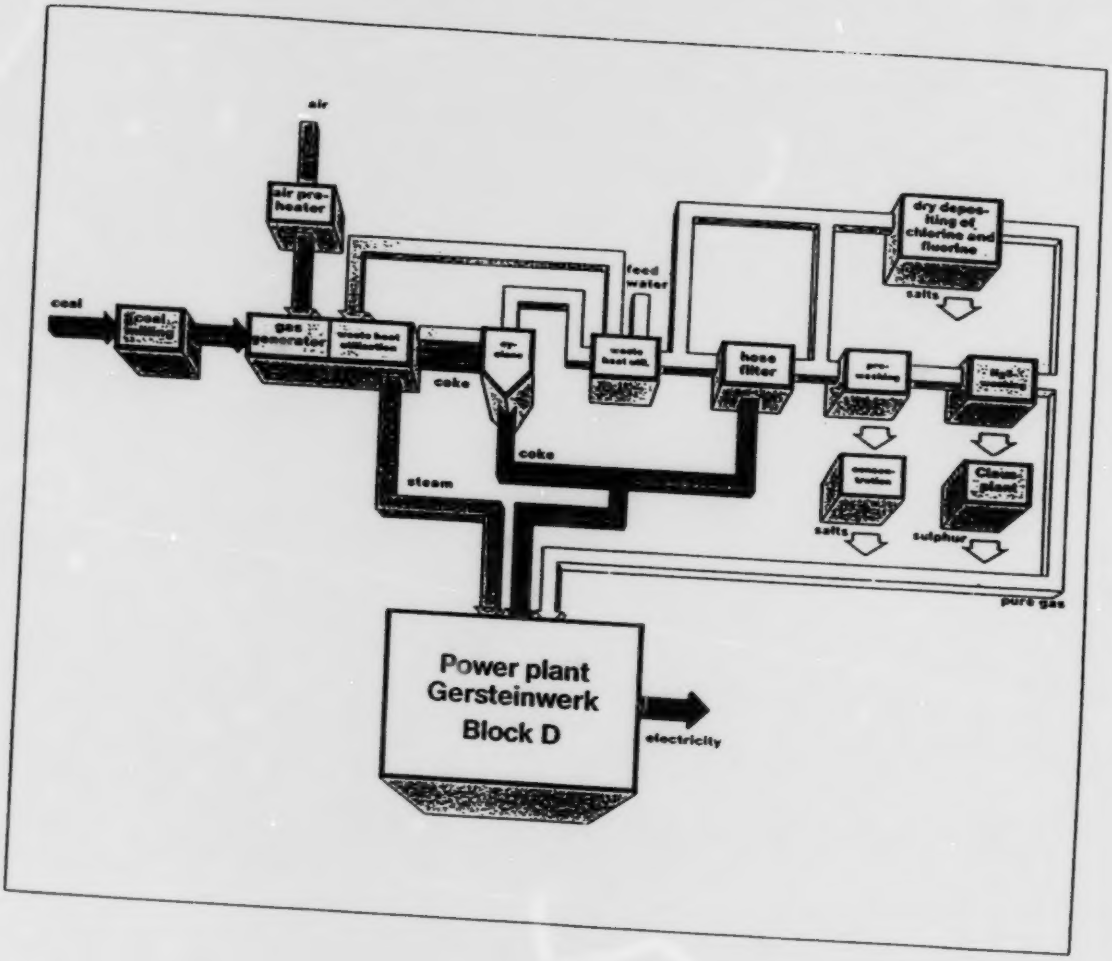

air
air pre-heater
coal
coal
gas generator
waste heat utilization
coke
coke
steam
feed water
waste heat util.
hose filter
dry depositing of chlorine and fluorine
salts
pre-washing
Claus-plant
salts
sulphur
pure gas
Power plant Gersteinwerk Block D
electricity

Figure 3.25

The Integration of the VEW Gasification Facility into the Electrical Generation Grid (VEW)

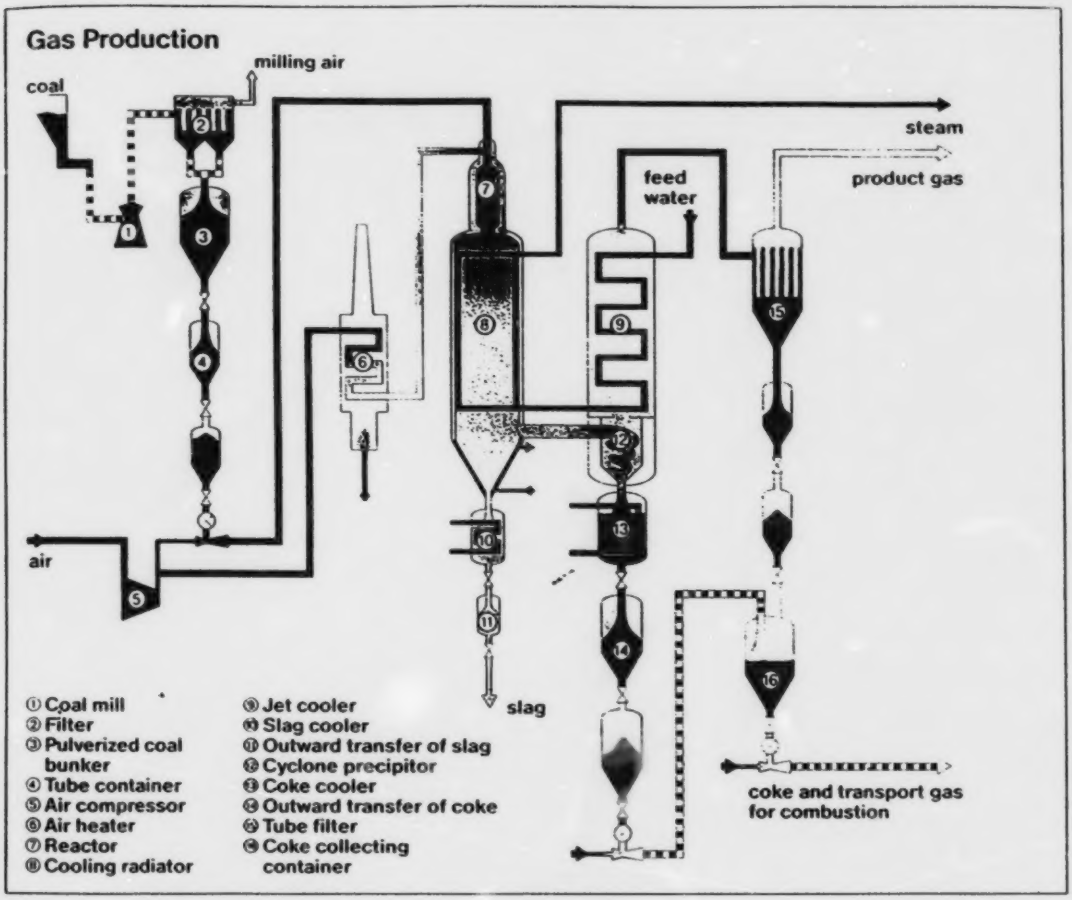

Gas Production
milling air
coal
steam
feed water
product gas
air
slag
coke and transport gas for combustion
① Coal mill
② Filter
③ Pulverized coal bunker
④ Tube container
⑤ Air compressor
⑥ Air heater
⑦ Reactor
⑧ Cooling radiator
⑨ Jet cooler
⑩ Slag cooler
⑪ Outward transfer of slag
⑫ Cyclone precipitor
⑬ Coke cooler
⑭ Outward transfer of coke
⑮ Tube filter
⑯ Coke collecting container

Figure 3.26

A Schematic of the VEW Gasifier System (VEW)

The reactor "burner" has variable radial and axial flow rate adjustments that allow the optimization of reactor performance. Control is exercised in order to maintain the lower heating value of the gas stream at approximately 4.1 MJ/Nm^3. The control of the gas heating value is valuable in optimizing the gas turbine performance.

A separate air stream enters the burner after it is preheated to 600°C in a preheater. The preheating improves the overall cycle efficiency. The present configuration uses a seven stage air compressor with intercooling. A commercial plant would use a compressor design that did not need intercooling and would utilize the heat gained during compression to eliminate the need for a separate air heater.

The refractory lined, 1 m ID reactor section is operated at 2 MPa and 1600°C. The gasifier converts 60 - 70% of the coal to gas with the remainder left as char. The reactor section is followed by a radiant cooler (Figure 3.26) that lowers the gas temperature to 1,300 - 1,400°C. Steam at 18.2 MPa and 535°C is generated in the radiant cooler and sent to the steam turbine. Approximately 10% of the ash in the coal leaves the reactor as slag (< 1% carbon composition). The dependence of the coal conversion on the air to coal ratio and the volatile content of the coal is shown in Figure 3.27. Both air to coal ratio and coal volatile content increases result in increases in the coal conversion in the gasifier.

A full scale reactor is considered to be in the 60 to 80 t/h coal throughput range. This would require an approximate $(70/10)^{0.5}$ ($\approx \times 2.65$) increase in the present reactor internal diameter.

After further cooling in the jet (convective) cooler, the coarse fraction of the char is separated in a cyclone at 800°C (Figure 3.26). The cooling in the reactor results in the need for less cooling in the jet cooler section. This contributes to a higher efficiency than that encountered in total gasification processes. The present configuration sees the char further cooled to 200°C and then fed to a pulverized coal designed boiler where electricity is generated. The low temperature energy is used in preheat applications. The commercial scale operation would send the 800°C char to a CFB.

The gas stream has Cl and F removed in a wet scrubbing system as shown in Figure 3.28. The Cl and F removal is carried out to eliminate the detrimental effect that these chemicals would have on the methanol-amine H_2S stripping process. A dry system that utilizes lime in a CFB at 2 MPa and 300°C has also been developed and used (Figure 3.29). The solid waste from the dry scrubbing system can be sent to the char combustion CFB to further enhance the utilization of the sorbent capacity of the lime. It is also possible to treat the adsorbent from the dry gas cleaning separately and to produce a dry salt.

A wet scrubbing system for SO_2 removal has also been operated (Figure 3.30). The stripped H_2S is then sent to a Claus plant to make sulphuric acid. A Claus plant does not operate well when following the load variations of an electrical generation station. The commercial application would not utilize the steps noted above but would send the gas stream to the CFB. The sulphur in the H_2S would eventually be reacted with CaO to form $CaSO_4$.

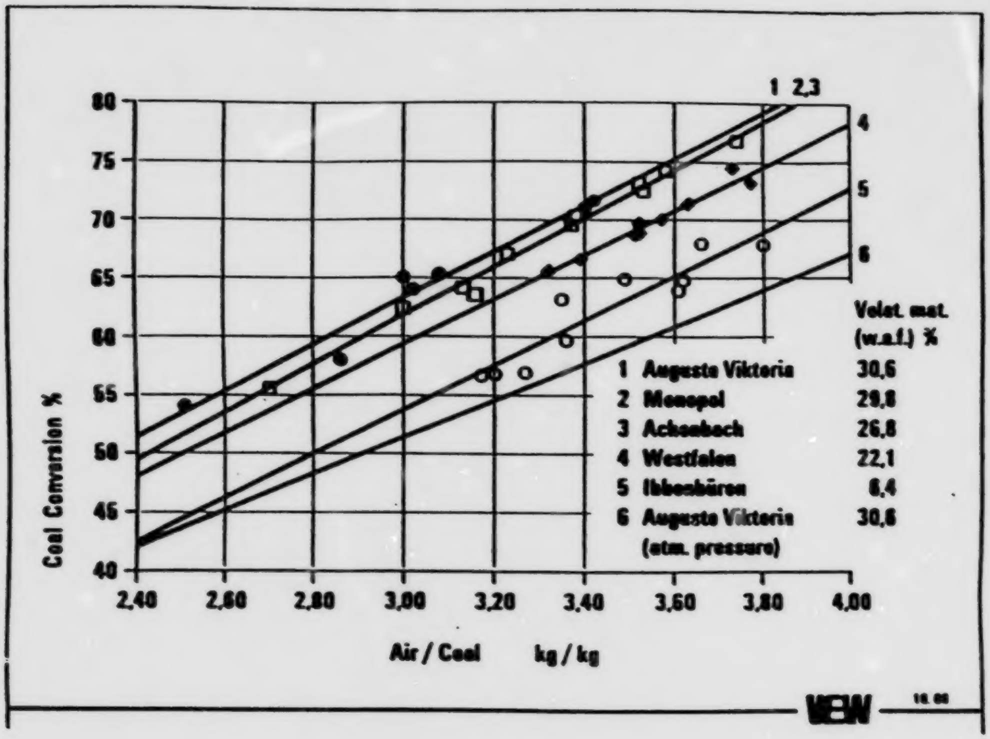


Figure 3.27

Effect of Air to Coal Ratio and Coal Volatile Content on the Conversion of the Coal in the VEW Gasifier

(Weinzierl, 1989)

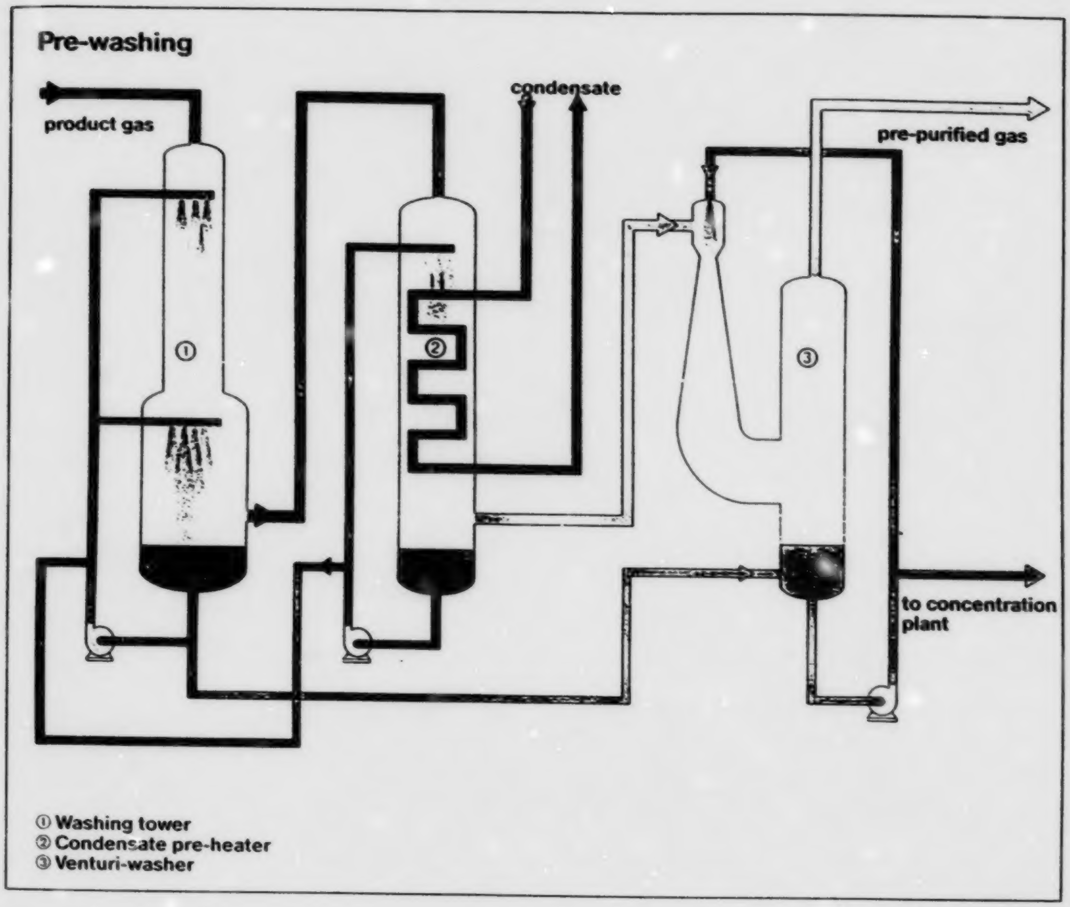

Pre-washing
product gas
condensate
pre-purified gas
①
②
③
to concentration plant
① Washing tower
② Condensate pre-heater
③ Venturi-washer

Figure 3.28

The VEW Chlorine and Fluorine Wet Scrubbing System (VEW)

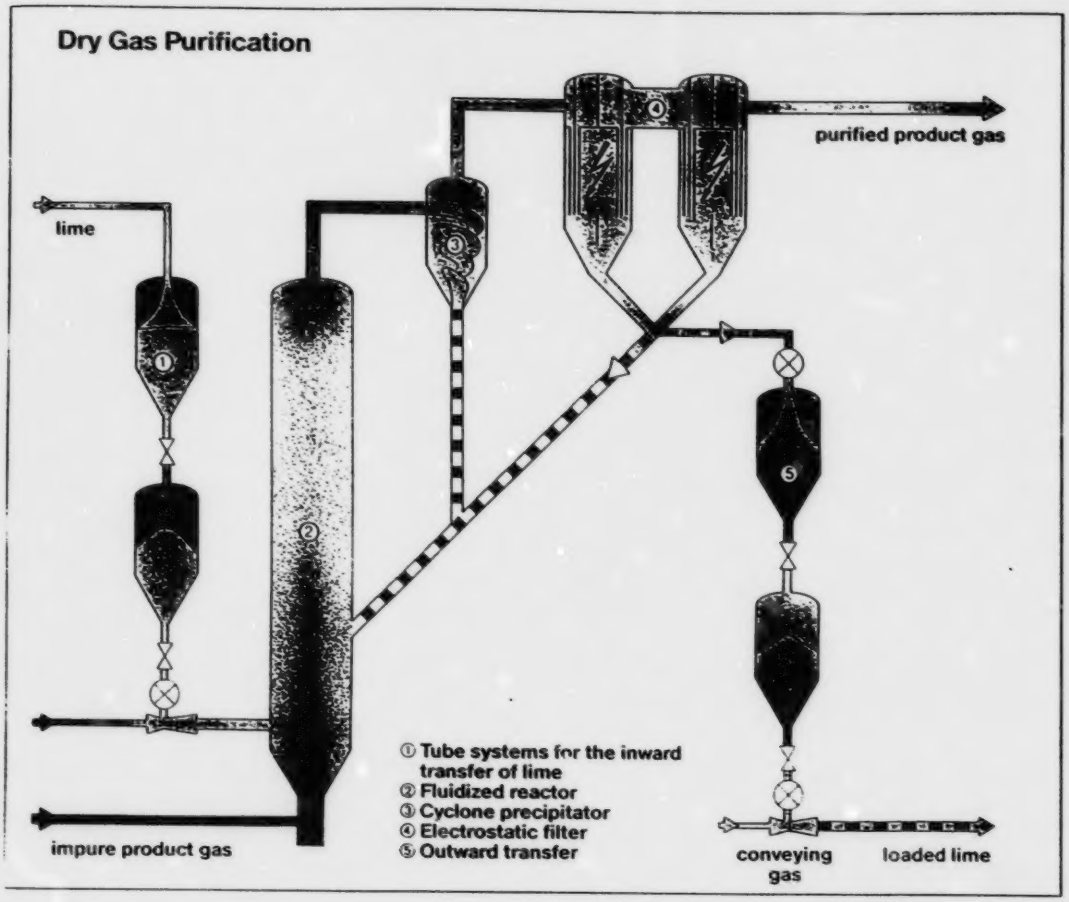

Dry Gas Purification
purified product gas
lime
④
③
①
②
⑤
impure product gas
① Tube systems for the inward transfer of lime
② Fluidized reactor
③ Cyclone precipitator
④ Electrostatic filter
⑤ Outward transfer
conveying gas
loaded lime

Figure 3.29

The VEW Chlorine and Fluorine Dry Scrubbing System (VEW)

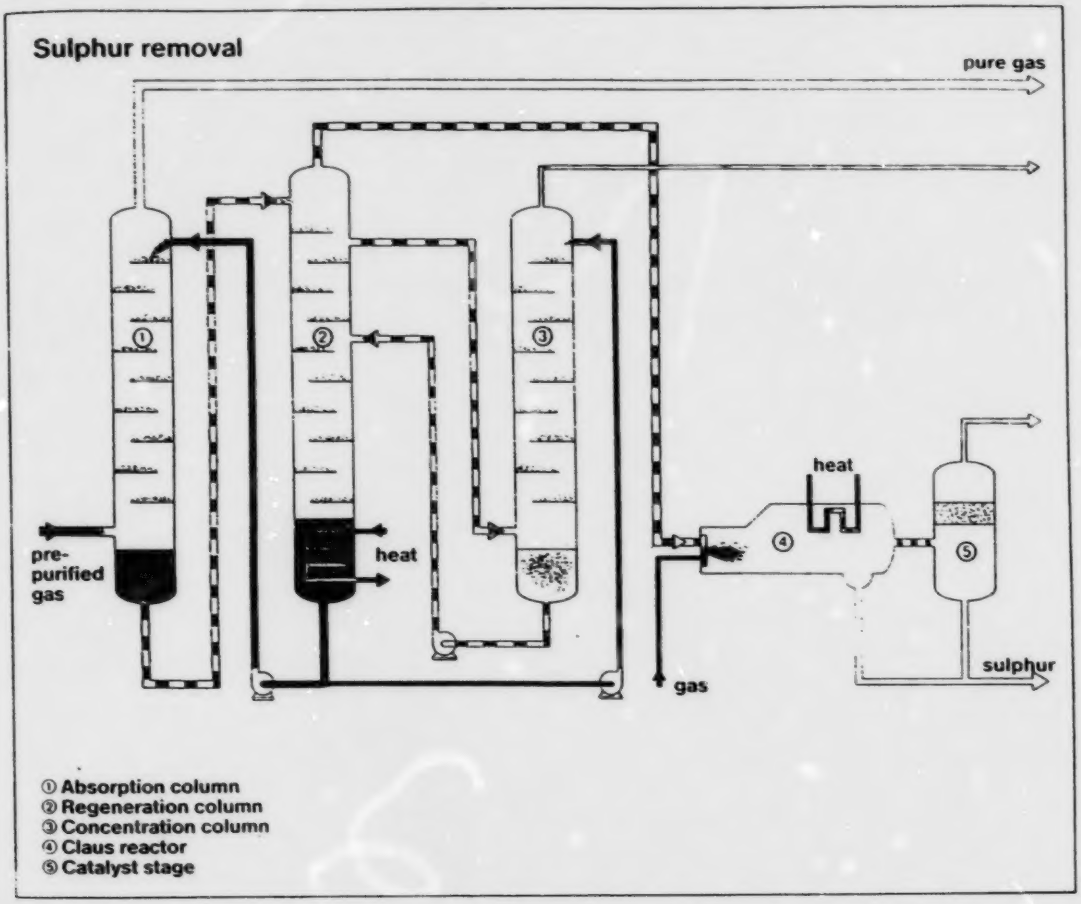

Sulphur removal
pure gas
heat
heat
pre-
purified
gas
gas
sulphur
① Absorption column
② Regeneration column
③ Concentration column
④ Claus reactor
⑤ Catalyst stage

Figure 3.30

The VEW SO_2 Wet Scrubbing System (VEW)

Char particles that are not separated in the char cyclone are presently removed in a bag house after the gas is cooled to 200°C (Figure 3.25). The energy removed from the gas stream is used to heat feed water and to generate steam. Testing is presently under way on electrostatic precipitators (ESP) that will replace the bag house and allow operation at higher temperatures (Figure 3.29). The ESP's are to be operated at 350°C and 2 MPa. The targeted particulate concentration is 10 mg/Nm^3 (N refers to 0°C and 101 kPa conditions). In a commercial operation VEW will utilize a particulate removal system that is optimum for the application (ESP, ceramic barrier filter, etc).

VEW has a test facility used to examine the corrosion behaviour of the gas turbine materials. Information from the corrosion studies will be used to optimize the operation of the plant and to minimize the blade deterioration.

In the present configuration, the cooled char is fed to a conventional pulverized coal fired design boiler (Figure 3.25). VEW is working with Steinmuller on the design of a CFB that would accept the hot char and exhaust from the gas turbine. Fluidizing and combustion O_2 will be from the turbine exhaust. A schematic of the proposed integrated gasification/CFBC system is shown in Figure 3.31. The CFB utilizes inertial vane separators near the top of the CFBC that eliminate the need for external hot cyclones (Figure 3.32). The hot char returns to the lower section of the CFBC along the walls. The char particles remain in an oxidizing atmosphere at high temperature which is beneficial in the low oxygen content gas from the gas turbine. The char produced in the gasifier is very reactive due to the high internal surface area (700 m^2/g) that is seen even in low volatile coals (i.e. 10% VM).

NO_x, NH_3 and HCN formed in the gas turbine are reacted in the CFB and therefore do not need to be removed prior to entering the CFB. The NO_x reduction mechanism is not understood. The CFBC char is suspected of acting as a catalyst in the reduction of the NO_x to N_2. This results in the NO_x concentration leaving the CFB that is virtually independent of the NO_x concentrations leaving the gas turbine. Limestone will be added to react with the sulphur containing species in the solid and gas streams entering the CFB. A Ca/S molar ratio of 2 is expected to be used to meet required emission restrictions in a CFB operated at 850°C.

VEW has 760 MWe of generating capacity based on cogeneration in operation at the same site as shown in Figure 3.33. Natural gas turbines (three) are exhausted into three conventional pulverized coal fired boilers. This information is valuable for the design of a coal fired gasifier combined cycle operation. The plant efficiency is reported (Weinzierl 1985) to be 42% and the availability similar to a conventional hard coal-fired power plant.

3.3.2.2 Review of Active Co-Product Producing Partial Gasification Processes

This section deals with the processes that are currently active and which produce co-products. Co-products include the solid, liquid and gas phase products that result from the partial gasification of coal. In particular, the solids could be char for combustion or the production of formed coke, the liquids could be a synthetic crude to be upgraded to diesel fuel or a gasoline additive and the gases typically would have a low heating value that would be best used in producing process heat.

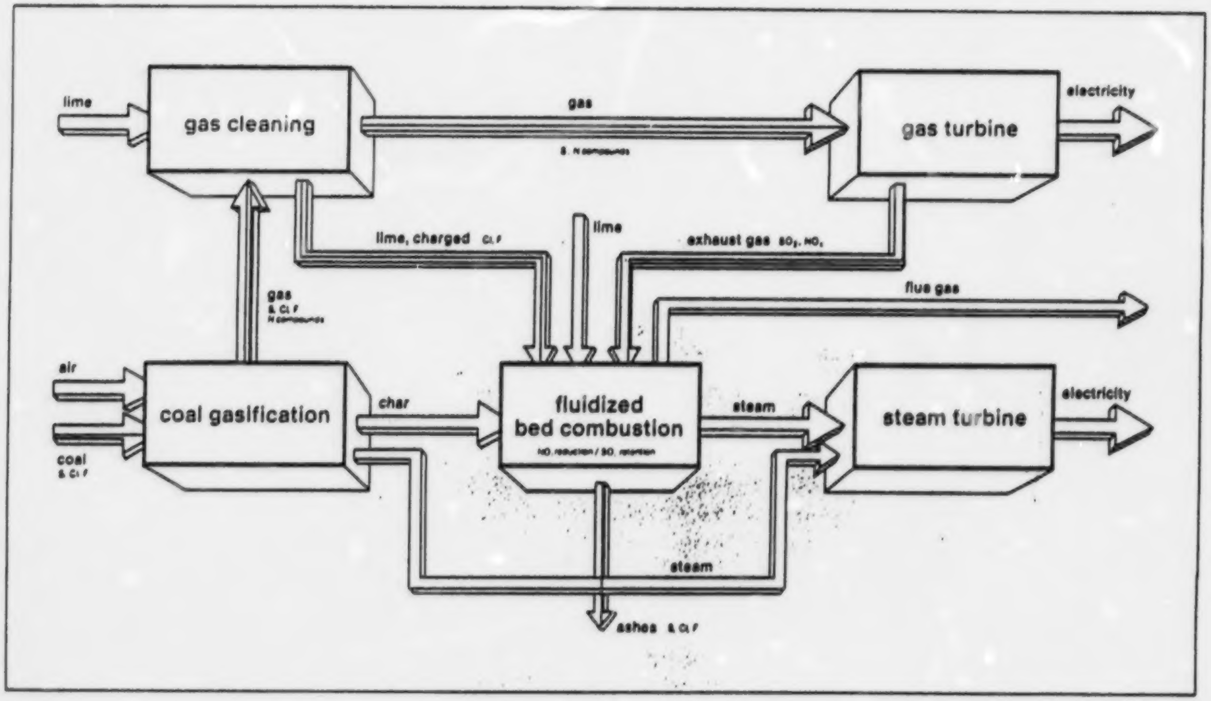

lime
gas cleaning
gas
S, N compounds
gas turbine
electricity
lime, charged Cl, F
lime
exhaust gas SO_2, NO_x
flue gas
gas
S, Cl, F
N compounds
air
coal
S, Cl, F
coal gasification
char
fluidized
bed combustion
NO_x reduction / SO_2 retention
steam
steam turbine
electricity
steam
ashes S, Cl, F

Figure 3.31

The VEW Gasifier/Circulating Fluidized Bed Combustor Concept (VEW)

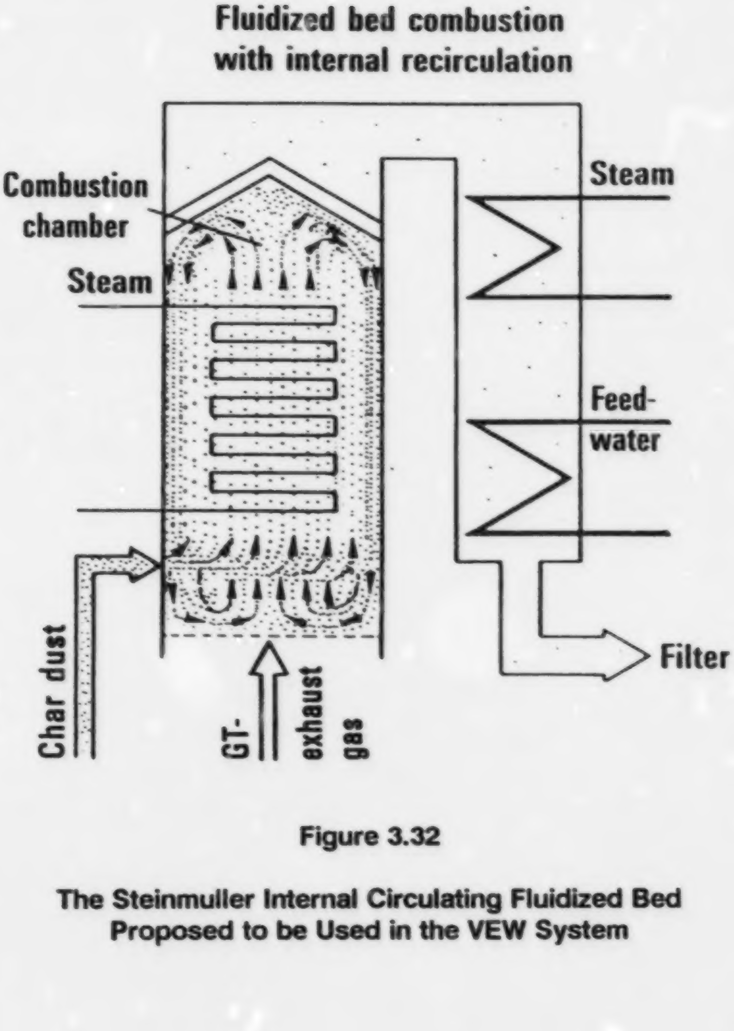
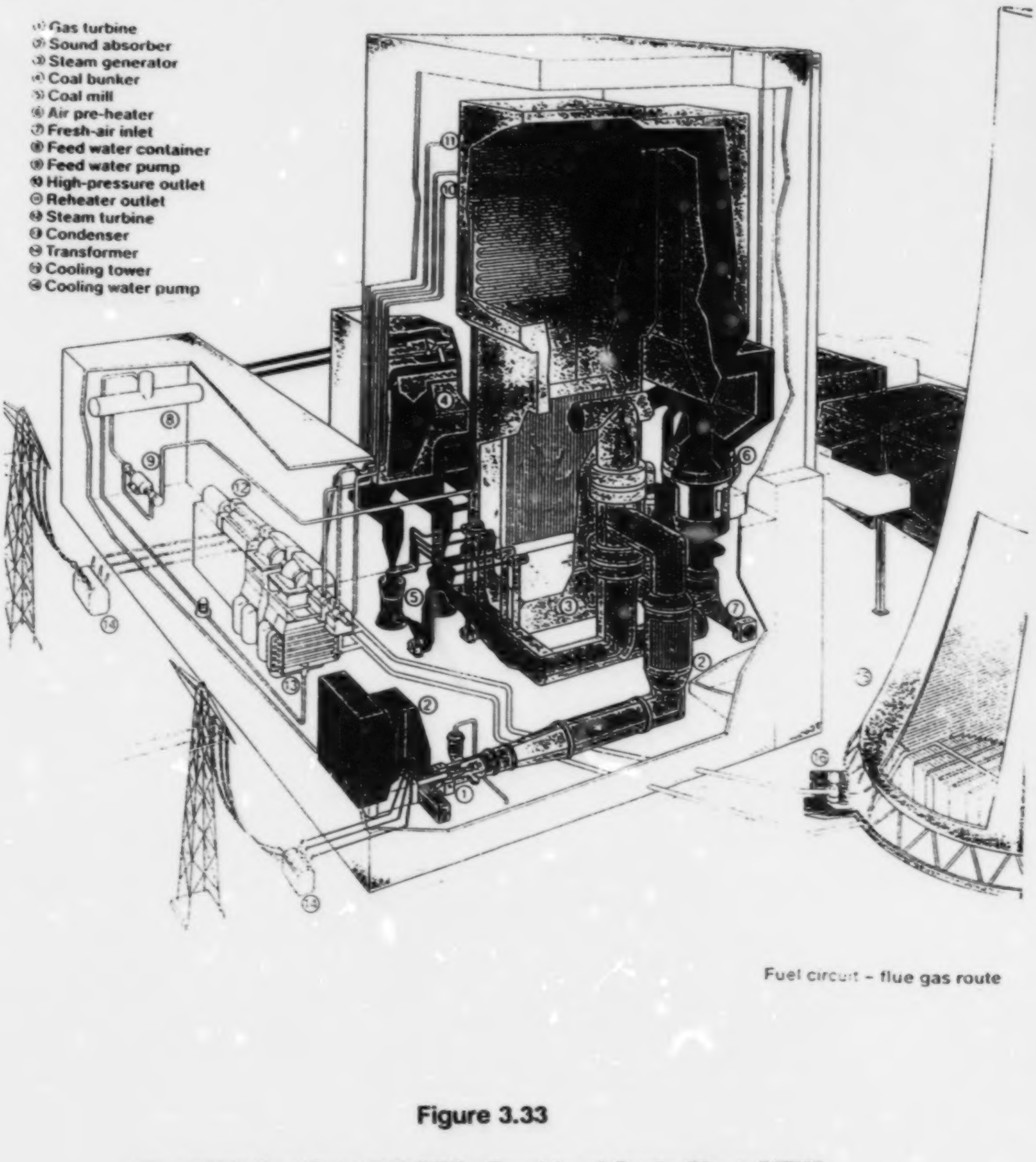

Fluidized bed combustion with internal recirculation
Combustion chamber
Steam
Steam
Feed-water
Filter
Char dust
GT-exhaust gas

Figure 3.32

The Steinmuller Internal Circulating Fluidized Bed Proposed to be Used in the VEW System



① Gas turbine
② Sound absorber
③ Steam generator
④ Coal bunker
⑤ Coal mill
⑥ Air pre-heater
⑦ Fresh-air inlet
⑧ Feed water container
⑨ Feed water pump
⑩ High-pressure outlet
⑪ Reheater outlet
⑫ Steam turbine
⑬ Condenser
⑭ Transformer
⑮ Cooling tower
⑯ Cooling water pump
①
②
③
④
⑤
⑥
⑦
⑧
⑨
⑩
⑪
⑫
⑬
⑭
⑮
⑯

Fuel circuit – flue gas route

Figure 3.33

The VEW Gas/Coal 750 MWe Combined Cycle Plant (VEW)

By far the most activity in this field is taking place in the United States. Table 3.31 shows the United States Department of Energy (USDOE) sponsored (August, 1990) coal gasification contractors.

Table 3.32 shows a summary of the USDOE funded partial gasification co-product processes.

British Coal

Gale et al. (1989) describes the development of the submerged spouted bed gasifier in two variants. One variant utilizes a coal derived char bed while converting approximately 60 to 70%, mass basis, of the coal into gas. The other variant uses lower coal to air mass flow ratios and an inert bed (i.e. Bauxite) to convert up to 90% of the coal to gas. The reactor is shown in Figure 3.19. The majority of the fluidizing gas is introduced as a jet at the apex of the conical base. Additional fluidizing gas can be injected in the conical section to ensure complete fluidization. Solids circulation caused because of the gas flow prevents agglomeration of caking coal particles.

Bed heights are used which are above the maximum spouted bed height, resulting in the upper section of the bed behaving in a slugging mode. Two coal feed sites may be used, either at the base of the cone or via a screw feeder above the bed.

Figure 3.34 shows a flow diagram of the British Coal 0.5 tonne/h Pilot Plant. The low conversion variant was operated for 700 hours. Table 3.33 shows results of this mode of operation. A representative analysis of the coal used is shown in Table 3.34. It should be noted that this system was designed to accept coal in the "singles" size range, that is 12 - 25 mm. Bed material was approximately 80% carbon. Recycling of fines that had been elutriated back to the bed resulted in an increase of only 2% (energy basis) in the conversion of coal to gas. Virtually no change was seen in the calorific value of the gas (3.1 MJ/m^3, net). Turndowns of up to 60% could be achieved.

Up to 80% sulphur capture was achieved using limestone utilizing a Ca/S molar ratio of 2 (Green et al., 1984).

As of the publication of this reference, a total of 1750 hours of testing had been carried out using a high conversion variant of the gasifier. An inert bed material of silica sand or Bauxite was used to reduce the carbon content of the bed to between 5 and 10%. This higher density Bauxite bed results in the need for more blower power. The use of silica sand as a bed material showed agglomeration as a result of interaction of iron oxide in the coal ash. Shallower bed depths of 1.5 m were used with the Bauxite bed which lowered the maximum coal conversion from 90% (1.5 m bed) to 86% (maf basis). The loss of conversion is more than offset by the reduction in the plant cost as the reactor size and weight is reduced. Turndowns of 45% full load have been demonstrated with no reduction in the heating value of the gas. Table 3.35 shows the relative performance data for sand and Bauxite beds with and without fines recycle. More dramatic improvements in the coal conversions as a result of fines recycle are seen in these data than that shown for the low conversion variant.

Table 3.31

United States Department of Energy Sponsored (August, 1990) Coal Gasification Contractors

GASIFICATION ACTIVITY	CONTRACTOR
Catalysis/H_2S Removal	Lawrence Berkeley Laboratory
Combustion	Advanced Fuel Research Inc. Donlee Technologies Inc. Brigham Young University
Electrode Carbon	AMAX R&D Center
Fuel Upgrade	Lummus Crest Inc.
Gas Cleaning/Sulphur Removal	Massachusetts Institute of Technology
Gas Component Separation	California Institute of Technology (H_2) CeraMem Corporation City College of New York (H_2) Energy Research Corporation (H_2) Louisiana State University (CO_2) MTCI (H_2) SRI International (syn gas)
Gas Cleaning/Particulate Removal	Air Products and Chemicals ALCOA Separation General Electric Corporate R&D Center Westinghouse S&T Center Worcester Ploytechnic Institute
Gasification	Brookhaven National Laboratory
Partial Gasification/Co-Product Production	Coal Technology Corporation ENCOAL Corp. Institute of Gas Technology Lawrence Livermore National Laboratory University of North Dakota Western Research Institute
Gasification/Power Generation/Gas Clean Cleaning	Southern Company Services
Gasification/Power Generation/Fuel Production	Calderon Energy Inc.
Gasification/Power Generation	Foster Wheeler General Electric Environmental Services M.W. Kellogg Company Pennsylvania Electric Company Southern Company Services Texaco Inc.
Partial Gasification/Power Generation	Foster Wheeler General Electric Environmental Services M.W. Kellogg Company Southern Company Services
Gasification/General	Brookhaven National Laboratory
Product Characterization	Oak Ridge National Laboratory
Synthetic Gas	SRI International
Product Toxicity	NIOSH
Various	Morgantown Energy Technology Center

Table 3.32

A Summary of USDOE Funded Co-Product Gasification Process

Process Name	Location, State	Scale (kg/h)	Reactor Design	Caking Coal
Coal Technology Corporation	Virginia	454	Screw	Yes
ENCOAL Corporation	Wyoming	41,000	Rotary Grate	No
Institute of Gas Technology	Illinois	45	Fluidized Bed/ Entrained Bed	Yes
University of North Dakota	North Dakota	45	Fluidized Bed	No
Western Research Institute	Wyoming	45	Fluidized Bed	No

Table 3.33

Performance of the British Coal 0.5 tonne/h Gasification Plant and Operating Conditions

(Gale et al., 1989)

Operating Conditions			
Mode	Without Fines Recycle		With Fines Recycle
Coal	Baddesley	Seamoor	Baddesley
Bed Material	Char	Char	Char
Fluidizing Gas			
- air, % vol	86	89	90
- steam, % vol	14	11	10
Fluidizing Velocity, m/s	2.0	2.0	2.0
Coal Feedrate, kg/h	490	470	450
Bed Height, m	2	2	2
Bed Temperature, °C	970	977	973
Process Performance			
Gas Composition			
CO_2, % vol (dry)	10.9	10.7	10.1
CO, % vol (dry)	13.1	13.7	14.5
H_2, % vol (dry)	14.3	14.5	13.2
CH_4, % vol (dry)	1.6	1.4	0.9
N_2, % vol (dry)	60.1	59.7	61.3
H_2O, % vol (dry)	14.0	11.2	8.9
Gas Calorific Value			
(Gross)$^{(1)}$, MJ/m^3	3.9	3.9	3.7
(Net)$^{(2)}$, MJ/m^3	3.1	3.2	3.1
Gas Output, m^3/h (wet)	1512	1470	1508
Coal Conversion Efficiency			
Mass Basis, %	65	63	69
Energy Basis, %	61	60	63

NOTES: (1) Gross = gross gas calorific value of dry gas calculated at 15°C and 1 atm
(2) Net = net gas calorific value of wet gas calculated at 15°C and 1 atm

Table 3.34

Representative Chemical Analysis of the Coal Used in the British Coal 0.5 tonne/h Gasifier Tests

(Gale et al., 1989)

Coal	Baddesley
Size, mm	12 - 25
Moisture, % ad	7.0
Ash, % ad	4.0
Volatile Matter, % ad	37.0
Carbon, % db	76.9
Hydrogen, % db	5.5
Nitrogen, % db	1.2
Sulphur, % db	2.0
Chlorine, % db	0.2
Calorific Value, MJ/kg (maf)	33.4
BS Swelling No.	1.0
Ash Deformation Temperature, °C	1190

Abbreviations:
ad = as determined
db = dry basis
daf = dry ash free

Table 3.35

Demonstration of Improved Coal Conversion Efficiencies on the British Coal 0.5 tonne/h Gasification Plant (Gale et al., 1989)

Operating Conditions				
Mode	Without Fines Recycle	With Fines Recycle	Without Fines Recycle	With Fines Recycle
Bed Material	Sand		Bauxite	
Fluidizing Velocity				
- air, % vol	100	95	90	90
- steam, % vol	0	5	10	10
Fluidizing Velocity, m/s	1.1	1.3	1.1	1.1
Coal Feedrate, kg/h	241	275	225	191
Bed Height, m	3.5	2.5	1.5	1.5
Bed Temperature, °C	1015	1020	1025	1020
Process Performance				
Gas Composition				
CO_2, % vol (dry)	6.6	7.3	11.2	10.2
CO, % vol (dry)	19.2	19.3	15.5	16.7
H_2, % vol (dry)	11.5	12.0	12.4	11.9
CH_4, % vol (dry)	1.0	0.7	1.0	0.6
N_2, % vol (dry)	61.7	60.7	60.1	60.6
H_2O, % vol (dry)	5.3	5.5	10.7	8.0
Gas Calorific Value				
(Gross)$^{(1)}$, MJ/m^3	4.1	4.0	3.9	3.8
(Net)$^{(2)}$, MJ/m^3	3.6	3.6	3.4	3.2
Gas Output, m^3/h (wet)	836	1114	813	885
Coal Conversion Efficiency				
Mass Basis, %	83	95	78	86
Energy Basis, %	79	91	74	80

NOTES: Baddesley Coal (12-25 mm) used in all tests, see Table 3.34 for representative chemical analysis.

(1) Gross = gross calorific value of dry gas calculated at 15°C and 1 atm.
(2) Net = net calorific value of wet gas calculated at 15°C and 1 atm.

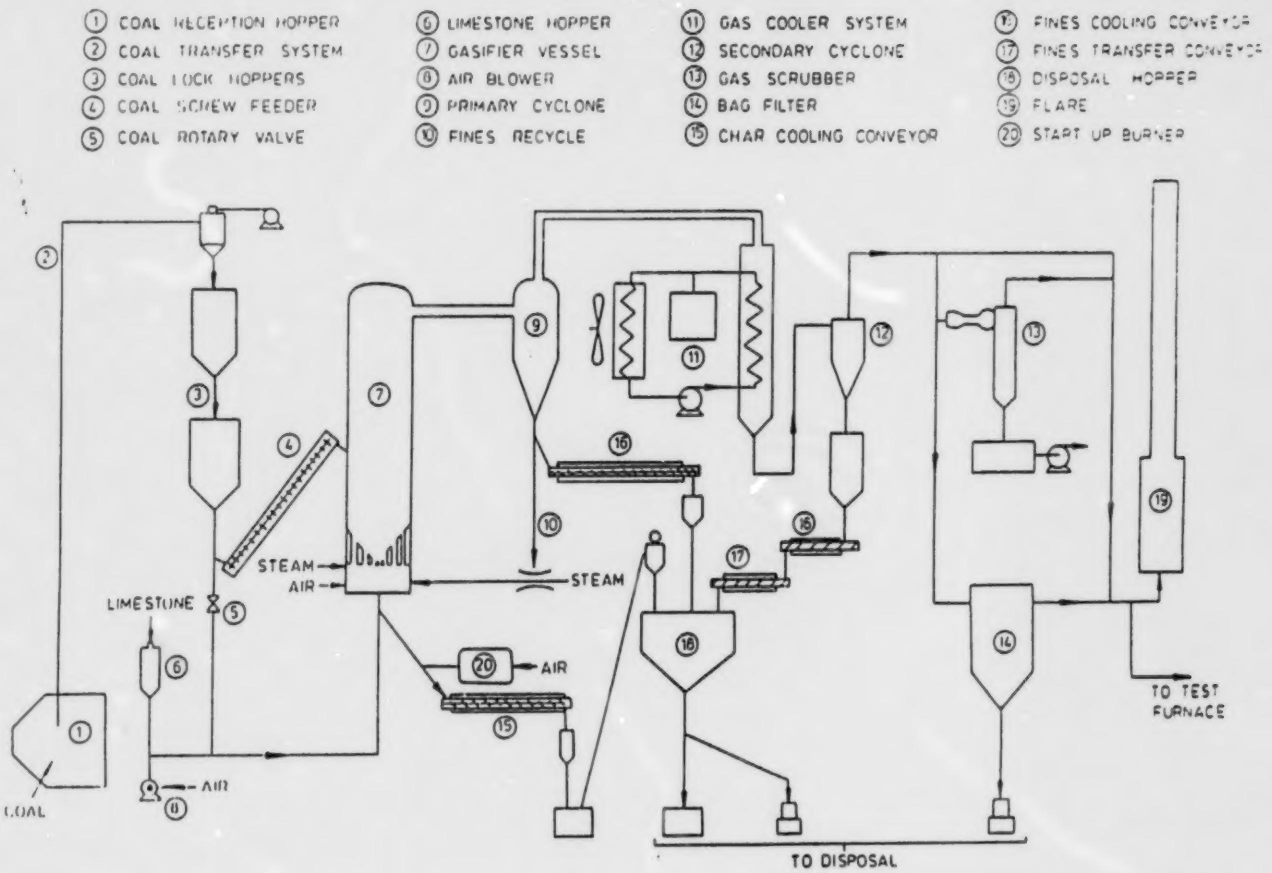

① COAL RECEPTION HOPPER
② COAL TRANSFER SYSTEM
③ COAL LOCK HOPPERS
④ COAL SCREW FEEDER
⑤ COAL ROTARY VALVE
⑥ LIMESTONE HOPPER
⑦ GASIFIER VESSEL
⑧ AIR BLOWER
⑨ PRIMARY CYCLONE
⑩ FINES RECYCLE
⑪ GAS COOLER SYSTEM
⑫ SECONDARY CYCLONE
⑬ GAS SCRUBBER
⑭ BAG FILTER
⑮ CHAR COOLING CONVEYOR
⑯ FINES COOLING CONVEYOR
⑰ FINES TRANSFER CONVEYOR
⑱ DISPOSAL HOPPER
⑲ FLARE
⑳ START UP BURNER
COAL
LIMESTONE
AIR
STEAM
AIR
STEAM
AIR
TO TEST FURNACE
TO DISPOSAL

Figure 3.34

The Cheltelham 500 kg/h Partial Gasification Facility As Built

(Gale et al., 1989)

Tests were carried out with enriched oxygen concentrations to determine if the calorific value of the gas could be increased. Table 3.36 shows the effect of increasing the oxygen concentration. The maximum gas calorific value obtained was 7.4 MJ/m^3 (gross) with 23% oxygen addition.

Comparative analysis of the combustion characteristics of the low calorific value gas (LCV) to natural gas and distillate fuel are shown in Table 3.37. Stable and efficient combustion of the LCV gas has been achieved using a redesigned commercial burner. British Coal succeeded in raising the flame temperature of the LCV in a CRE test furnace by:

- preheating the combustion air to 505°C and the LCV gas to 360°C
- enriching the oxygen concentration to 36% (by volume)
- raising the calorific value of the gas to 7 MJ/m^3

Table 3.38 shows a summary of emissions to air from the variants of the 500 kg/h process.

The calcium sulphide in the solid waste streams is easily oxidized to calcium sulphate under mild oxidation conditions in a reactor. H_2S is not emitted if the residues are not exposed to strongly acidic leachates.

Coal Technology Corporation Process

The design philosophy that the Coal Technology Corporation (CTC) (Chu et al., 1989) has used is to convert coal to the highest-value-added products in a fashion that minimizes the plant capital costs in order to optimize the overall economics of the process. The co-products that have been targeted are char and coal liquids.

Data have been generated using three reactor systems to allow scale up to a product development unit (approximately 500 kg/h). The three units are a batch bench scale unit called the MGU (Figure 3.35), a 7.6 x 7.6 cm batch unit (BSR) (Figure 3.36) used to simulate a block of coal in a 1 t/h unit and a 3.8 cm diameter single screw unit (Figure 3.37) operated at Lawrence Livermore National Laboratory. In addition, liquid upgrading and utilization tests have been carried out.

The bench scale mild gasification (MGU) unit (Figure 3.35) was used to carry out a parametric testing program. This resulted in the establishment of optimum operating conditions. These conditions were not reported (Chu et al., 1989). The bench scale unit is a batch process with an off gas condensing system.

The BSR unit shown in Figure 3.36 was used to simulate one block of the coal bed in a 500 kg/h scale reactor. This unit is used to produce the char that is used to make the formed coke. No data are reported on the operation of this unit.

Table 3.36

Performance of the British Coal 0.5 tonne/h Gasifier with Oxygen Enriched Fluidizing Gases (Gale et al., 1989)

Operating Conditions			
Primary Fines Recycle	No	No	No
Fluidizing Gas			
- added oxygen, % vol	0	12	23
- air, % vol	90	62	52
- steam, % vol	10	26	25
- total oxygen, % vol	17	24	34
Fluidizing Velocity, m/s	1.1	1.1	1.1
Coal Feedrate, kg/h	225	280	451
Bed Height, m	1.5	1.5	1.5
Bed Temperature, °C	1015	1020	1020
Process Performance			
Gas Composition			
CO_2, % vol (dry)	16.2	14.6	14.5
CO, % vol (dry)	15.5	23.9	28.8
H_2, % vol (dry)	12.4	24.4	26.8
CH_4, % vol (dry)	1.0	1.1	1.3
N_2, % vol (dry)	60.6	36.0	28.6
H_2O, % vol (dry)	10.7	15.0	14.8
Gas Calorific Value			
(Gross)$^{(1)}$, MJ/m^3	3.9	6.5	7.4
(Net)$^{(2)}$, MJ/m^3	3.1	4.9	5.7
Gas Output, m^3/h (wet)	813	828	968
Potential Heat Output, MW (Net)	0.70	1.13	1.53
Coal Conversion Efficiency			
Mass Basis, %	78	78	82
Energy Basis, %	73	73	79

NOTES: Baddesley coal (12-25 mm) used in all tests, see Table 3.34 for representative chemical analysis.

(1) Gross = gross calorific value of dry gas calculated at 15°C and 1 atm
(2) Net = net calorific value of wet gas calculated at 15°C and 1 atm

Table 3.37

Combustion Properties of Low Calorific Value Fuel Gas From the British Coal 0.5 tonne/h Gasifier Compared with Natural Gas and Distillate Fuel

(Gale et al., 1989)

Fuel	LCV Gas(1)	Natural Gas	Distillate Fuel
Fuel Volumetric Flow Rate(2), $m_n{}^3/h$(6)	101	8.47	0.0088
Combustion Air Flow Rate(2), $m_n{}^3/h$(6)	115	112	112
Stoichiometric Adiabatic Flame Temperature, °C	1390(3)	1950(4)	2028(4)
Laminar Flame Speed(7), (% of speed of H_2 in air)	8.9	14.0	-
Emissivity	0.22(3)	0.22(3)	0.70(5)

NOTES:
(1) 3.8 MJ/m^3(6) (Gross) as in Table 3.35 (Bauxite bed with fines recycle)
(2) Calculated using (1) and (4)
(3) "Production of Industrial Fuel Gas for Applications Requiring High Flame Temperature and High Emissivity" - final report on ECSC Project No. 7220-EC/832
(4) "Combustion Engineering & Gas Utilization" - British Gas School of Fuel Management
(5) "Energy Managers Handbook", National Industrial Fuel Efficiency Service Ltd., 1985
(6) m^3/h = gas volume at 15°C and 1 atm
(7) US Bureau of Standards Journal of Research, 46, 211, 1951

Basis
100 kW
5%O_2 in flue gas

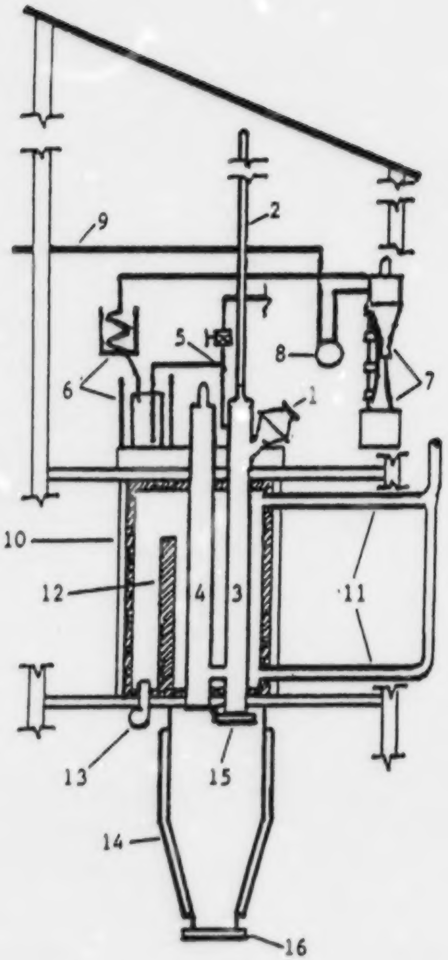
Table 3.38

Summary of Emissions to Air from the British Coal 0.5 tonne/h Gasifier

(Gale et al., 1989)

Species	Emission Level		Comments
	Moderate Conversion Variant	High Conversion Variant	
Dust, $mg/m_n^{3(1)}$	390 - 960	300 - 750	Two single cyclonic clean-up
	< 10	< 10	Two stage cyclonic clean-up + bag filter
Sulphur Dioxide ($6\% \ O_2$, dry) ppm	900	1200	Without limestone
	20 - 300	150	With limestone
Nitrogen Oxides ($6\% \ O_2$, dry) ppm	135	140	

Notes: (1) m_n^3 = gas volume at 15°C and 1 atm



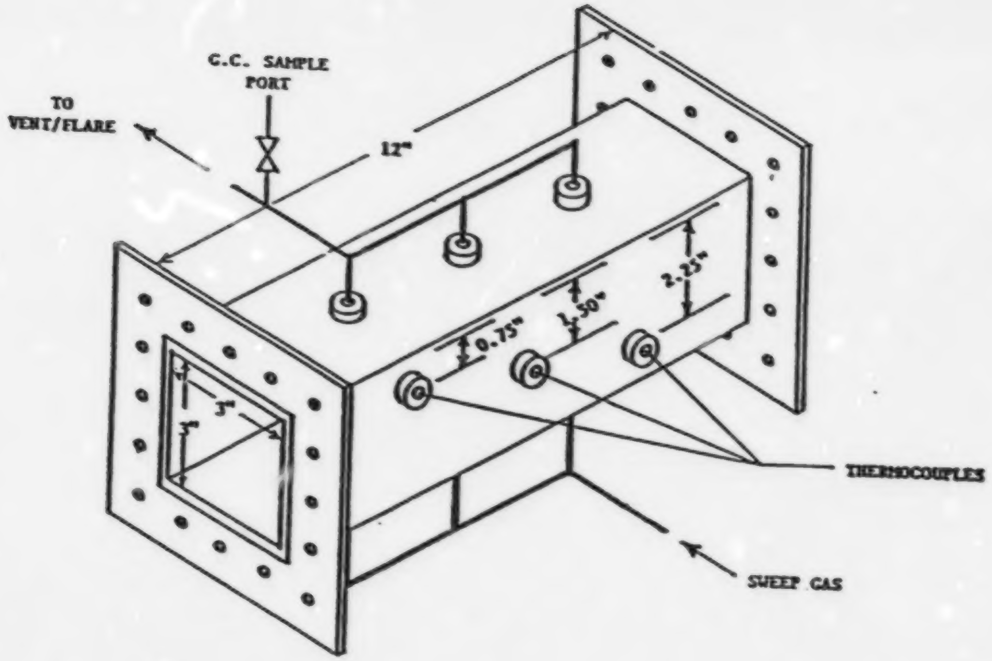
LIST OF COMPONENTS

1	VOLUMETRIC COAL FEED HOPPER (2)
2	HYDRAULIC CYLINDER FOR PLUNGER (2)
3	6" STAINLESS STEEL REACTOR TUBE (2)
4	8" STAINLESS STEEL SWEEP GAS HEATER TUBE (2)
5	GAS EXIT PIPE
6	SLIP-STREAM CONDENSING SYSTEM
7	CYCLONE, VENTURI-SCRUBBER, & TRAY TOWER CONDENSING SYSTEM
8	NON-CONDENSIBLE GAS VACUUM PUMP
9	NON-CONDENSIBLE GAS FLARE
10	FURNACE
11	FURNACE STACKS
12	FIRE BRICK BAFFLE
13	NATURAL GAS BURNER
14	CHAR COOLING UNIT w/ WATER JACKET
15	REACTOR TUBE DISCHARGE GATE
16	CHAR HOPPER DISCHARGE GATE

Figure 3.35

The CTC Bench Scale Mild Gasification Unit (MGU)

(Chu et al., 1989)



C.C. SAMPLE PORT
TO VENT/FLARE
12"
2.25"
1.50"
0.75"
3"
3"
THERMOCOUPLES
SWEEP GAS

Figure 3.36

The CTC 1.6 kg Bench Scale Reactor (BSR)

(Chu et al., 1989)

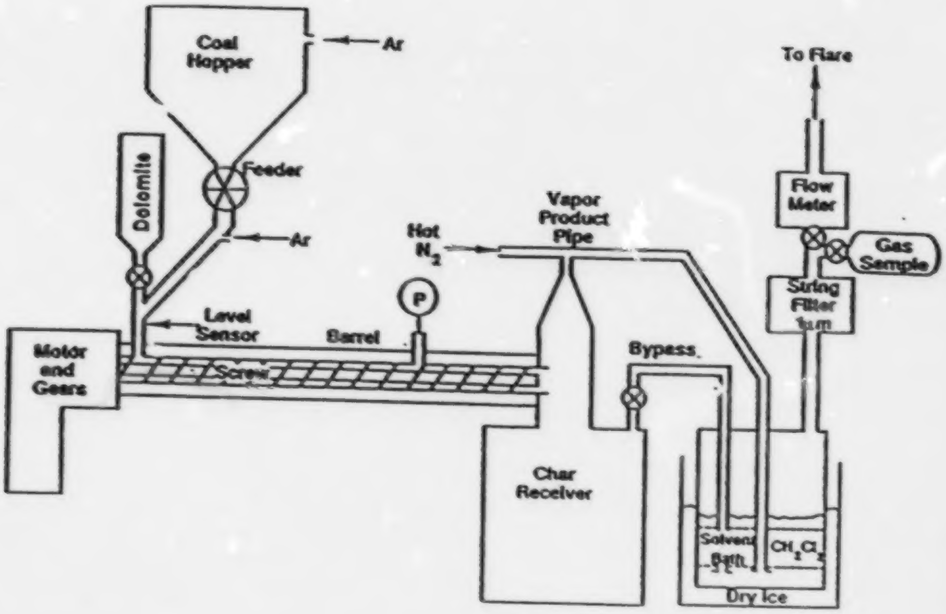

Coal Hopper
Ar
To Flare
Dolomite
Feeder
Ar
Vapor Product Pipe
Hot N_2
Flow Meter
Gas Sample
String Filter 1μm
P
Level Sensor
Barrel
Motor and Gears
Screw
Bypass
Char Receiver
Solvent Bath
-3°C
CH_2Cl_2
Dry Ice

Figure 3.37

The CTC LLNL 38 mm Single-Screw Pyrolyzer System

(Camp, 1990)

Camp et al. (1989) at the Lawrence Livermore National Laboratory, has worked in conjunction with CTC to develop a screw type reactor to carry out partial gasification (Figure 3.37). The single screw reactor is designed to operate on caking bituminous coals and is externally heated. Camp (1990) recommended that three processes be considered for further U.S. development for the continuous partial gasification of caking coals. They were the Occidental-hot-recycle-char entrained-flow flash pyrolyzer, the Lurgi-Ruhrgas hot-recycle-char screw-mixer flash pyrolyzer and the externally-heated screw pyrolyzer (both single and double screw versions). The externally heated reactor was chosen because it did not have recirculating solids to handle and separate and it uses little sweep gas, resulting in lower gas handling costs. The disadvantage is that the use of a mechanical device to control plastic reacting coal is unproven. Three screw designs were considered; 1) a single screw, weld fabricated, 2) twin screw, weld fabricated and 3) twin screw, machined. The three design options are shown in Figure 3.38. The reliability increases with the cost from option 1 to 2 to 3. Emphasis has been put on weld fabricated screw options due to the lower production costs.

Figure 3.39 shows the various stages that the coal would go through in the screw type reactor. The morphology of the coal varies as the temperature increases. The coal remains as particulate and changes to plastic as the softening point is reached. After the volatiles release stage, the remaining solids are friable char. The temperature of the coal increases until the mid section of the volatile collection zone and remains constant until it exits the reactor. A commercial system with a separate drying step would see the coal enter the reactor in a preheated state.

Tests were carried out on an East Kentucky coal having the analysis shown in Table 3.39.

Table 3.39

Analysis of the East Kentucky Coal Used in the LLNL Reactor

FSI	6.5
Volatiles (dry basis)	35.1%
Ash	4.1%
Sulphur	1.0%
H/C ratio	0.75
HHV	33.44 MJ/kg
Size	< 3 mm

The reactor used had a screw that was 37.9 mm OD and 89 cm long. The clearance between the screw and barrel is 0.3 mm. Two variations of the single screw design were tried (specifics of the design are not given).

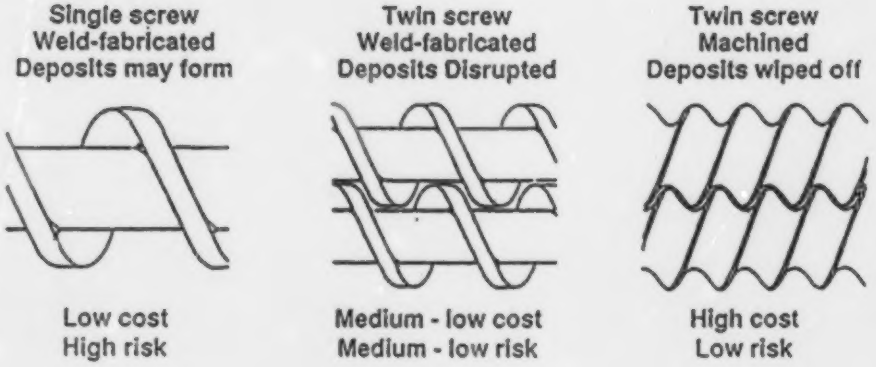
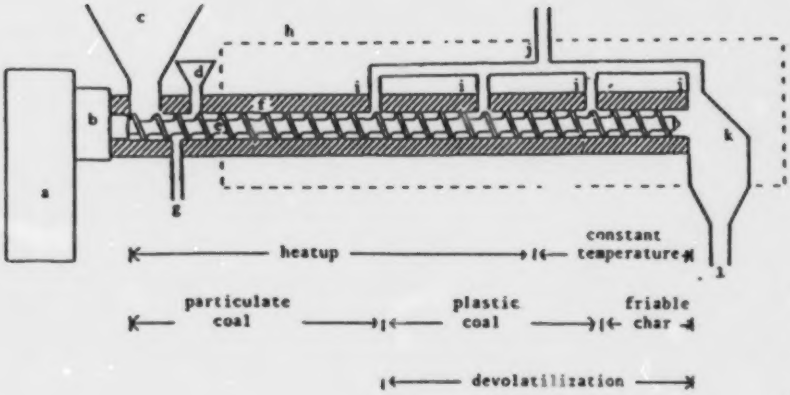

Single screw
Weld-fabricated
Deposits may form
Twin screw
Weld-fabricated
Deposits Disrupted
Twin screw
Machined
Deposits wiped off
Low cost
High risk
Medium - low cost
Medium - low risk
High cost
Low risk

Figure 3.38

Screw Design Options for the CTC Reactor

(Camp, 1990)


c
h
j
d
i
i
j
j
f
b
e
k
a
g
heatup
constant temperature
l
particulate coal
plastic coal
friable char
devolatilization

a. motor and gears

b. bearings

c. coal feed hopper

d. second solids hopper (optional)

e. screw shafts and flights (threads)

f. barrel

g. gas injection port (optional)

h. surrounding furnace chamber provides heat (could be replaced by electrical band heaters)

i. devolatilization ports

j. product plenum and line to condenser

k. char deflector and collector

l. char outlet

Figure 3.39

Coal Reaction Stages in a Screw Reactor

(Camp, 1989)

The following problems occurred frequently:

- plugging of the vapour removal ports, pipes and pressure taps by plastic coal, char fines, coke and tar

- back flushing of oil vapour that condenses and gums things up near the inlet

- high torque on the screw

- difficulty in removing the screw from the barrel after the runs

The maximum coal throughput is a function of the screw design and asymptotically reaches a maximum at low screw rpm for both designs (125 and 60 g/min for different unspecified screw designs).

The major problems encountered were that high throughputs could not be achieved and that the power requirements for screw rotation were high. Power to turn the screw was erratic and ranged from 100 to 900 J/g coal. This was a function of the rpm and the degree of fullness of the screw. It is noted that 775 J/g is needed to heat coal from 100 to 500°C.

Coating the screw material did not reduce torque requirements. It is hypothesized that a layer of coke deposits on the screw mitigating any effect of the screw material on torque requirements. Further screw design and scale up work is to be carried out.

A twin screw design is expected to have fewer operational problems. The interaction of the two screws should shear any adhered coal from the coal surface.

Laucks (1927) describes the development and operation of single vertical screw carbonizers. The most significant modification that was made to eliminate the caking of coal on the retort was control of the surface temperature. The original apparatus was externally heated. This left the screw surface temperature much less than that of the outer reactor tube wall. It was found that the coal that was made plastic from the heat from the outer tube wall would adhere to the lower temperature screw surface. Subsequent reactors were designed with heating applied both internally and externally. Years of operation were carried out with a 31 cm ID reactor. A 91 cm ID unit was constructed and operated better than the 31 cm reactor. At the time of the writing of Laucks' paper, a 2.7 m ID reactor was under construction. The 91 cm reactor had a capacity of approximately 500 kg/h. The primary product was smokeless fuel.

Variations in the rotation speed of one of the screws would further increase the shearing of the adhered coke on the screw surface.

Modelling of the heat transfer characteristics and the empirical data generated at LLNL resulted in the generation of an equation that predicts mass throughput of coal as a function of various parameters. The model suggests that the screw should be as long as is mechanically practical. This result has led to the design of a reactor system with a separate dryer in series with the reactor (see Figure 3.40). This configuration has a long effective screw length. In addition, optimization of conditions may be achieved and removal of water is in a separate step from the hydrocarbon evolution. This simplifies the product separation required. The model also suggests that the screw diameter should be as small as possible and that the rate of rotation should be high.

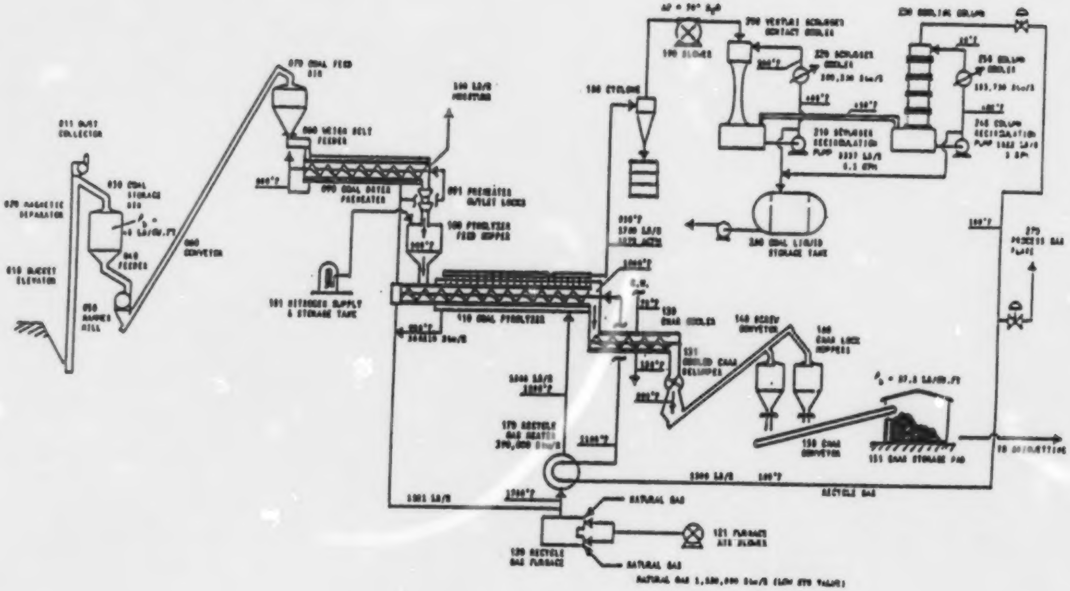

011 DUST COLLECTOR
020 MAGNETIC SEPARATOR
010 BUCKET ELEVATOR
030 COAL STORAGE BIN
040 FEEDER
050 HAMMER MILL
060 CONVEYOR
070 COAL FEED BIN
080 WEIGH BELT FEEDER
090 COAL DRYER PREHEATER
091 PREHEATER OUTLET LOCKS
100 PYROLYZER FEED HOPPER
101 NITROGEN SUPPLY & STORAGE TANK
110 COAL PYROLYZER
120 RECYCLE GAS FURNACE
121 FURNACE AIR BLOWER
130 CHAR COOLER
131 COOLED CHAR DELUMPER
140 SCREW CONVEYOR
150 CHAR CONVEYOR
151 CHAR STORAGE PAD
160 CHAR LOCK HOPPERS
170 RECYCLE GAS HEATER
180 CYCLONE
190 BLOWER
200 VENTURI SCRUBBER CONTACT COOLER
210 SCRUBBER RECIRCULATION PUMP
220 SCRUBBER COOLER
230 COOLING COLUMN
240 COLUMN RECIRCULATION PUMP
250 COLUMN COOLER
260 COAL LIQUID STORAGE TANK
270 PROCESS GAS FLARE
NATURAL GAS
NATURAL GAS
RECYCLE GAS

Figure 3.40

The CTC Proposed Process Development Unit

(Chu et al., 1989)

Char used to make coke is from the CTC BSR system. The CTC char coking facility is shown in Figure 3.41. Temperatures of 1010°C can be achieved. CTC leased a briquetting machine to produce the formed coke.

A coke reactivity index (CRI) is determined by reacting 200 g of 1.9 x 2.2 cm dry coke with CO_2 flowing at 5 L/h for 2 hours at 1100°C. The CRI value is the percent weight loss of the sample after the reaction. The target value of the CRI is < 30%. A coke strength after reaction (CSR) is determined by tumbling the coke from the CRI test in a drum for 600 revolutions at 20 rpm. The cumulative percent of the coke > 0.95 cm after the tumbling is reported as the CSR. The target value of the CSR is > 55 %. These are not the industry standard test procedures which means that the CRI and CSR values cannot be compared to other test results. The modified procedures are used to reduce the testing time. Table 3.40 shows coke test results. This data indicates that approximately 65% char, 25% binder coal and 5% coal tar produces an acceptable product. The values of variables associated with the production of the coke (such as pressure, temperature, particle size) are not given.

CTC has reported the following from the operation of the char-coker (Chu et al., 1989):

- modifications are needed in the char composition in order to make coke
- briquetting of the char prior to coking enhances the char coking process
- binder quality and content, briquetting pressure and temperature, blending of coking coal as a binder agent during coking, char and coke particle size and the total volatile content of the blend are all important factors in the coke production.

In order to gain experience in the upgrading of coal derived liquids prior to generation of their own, CTC (Miin et al., 1989) purchased 76,000 L of coal derived liquids from Coalite Ltd, Chesterfield, England. The Coalite process uses a vertical fixed bed reactor which is externally heated and has approximately 10 wt % yield of tar/oil. The ultimate analysis of the Coalite liquid is shown in Table 3.41.

Table 3.42 shows the ultimate analysis and the distillation data for the CTC Pro-blend and the Coalite coal-derived liquid. Figure 3.42 shows the distillation curves for the two liquids. This data can be used for superficial comparative purposes only, due to the differences in the coal feedstocks and the processes used to generate the liquids.

CTC has purchased and erected a 160,000 L/d coal liquid refinery to upgrade the coal derived liquids. A flow sheet of the refinery is shown in Figure 3.43. The refinery produces three fuel products simultaneously from coal liquids, diesel fuel, gasoline blending stock and heavy industrial fuel.

The Coalite liquid was upgraded as in the process shown in Figure 3.43. The colour of the coal-derived diesel darkened faster than petroleum derived diesel. The coal-derived fuel has less long term precipitation than petroleum fuel. Conventional stabilizers failed to affect the coal-derived liquids. CTC notes this is an area that needs further development.

The resultant coal-derived diesel fuel had a distillation curve (Figure 3.44) that closely resembled that of locomotive and commercial diesel fuels. Table 3.43 shows a comparison of commercial and CTC-diesel No. 2.

Table 3.40

Summary of Results on Selected Successful CTC Briquette Tests

(Im et al., 1990)

Date	Briquette No.	Char	Binder Coal	Hard Pitch	Other Binder Agents	CRI	CSR	Note
4/25/90	14	58.0%	27.0%	0%	9% Sodium Silicate 6% Corn Starch	46.2	12.8	Note 1
5/22/90	15	65.0%	25.0%	10%		26.0	77.7	
5/22/90	16		25.0%	10%	65% Devolatilized Char	39.6	26.9	
6/04/90	30	65.0%	25.0%	10%		47.7	28.7	
6/04/90	31	65.0%	25.0%	10%		33.8	69.2	
6/04/90	32	65.0%	25.0%	5%	5% Coal Tar	28.1	64.2	
6/04/90	33	65.0%	25.0%	5%	5% Coal Tar	19.4	80.9	
6/07/90	34	65.0%	25.0%	10%		21.1	77.8	
6/08/90	35	65.0%	25.0%	5%	5% Coal Tar	13.3	84.4	
6/13/90	36	65.0%	25.0%	5%	5% Coal Tar	31.0	64.0	Note 1
6/14/90	37	90.0%	0.0%	0%	10% Resin	17.1	27.0	Note 2
6/18/90	39	65.0%	25.0%	5%	5% Coal	29.7	49.4	Note 3
6/20/90	40	70.0%	30.0%	0%		24.8	50.9	
6/26/90	41a	65.0%	25.0%	5%	5% Coal Tar	17.5	85.9	
6/26/90	41b	65.0%	25.0%	5%	5% Coal Tar	34.3	57.4	Note 1

Table 3.40 (Continued)

Summary of Results on Selected Successful CTC Briquette Tests

(Im et al., 1990)

Date	Briquette No.	Char	Binder Coal	Hard Pitch	Other Binder Agents	CRI	CSR	Note
6/27/90	42	65.0%	25.0%	5%	5% Coal Tar	17.0	85.3	
7/06/90	45	65.0%	25.0%	5%	5% Coal Tar, Coalite char	35.8	73.0	
7/09/90	46	65.0%	30.0%	5%	Coalite Char	25.0	79.3	
7/24/90	48	65.0%	25.0%	0%	5% Type B pitch 5% Coal Tar	39.0	36.4	
7/30/90	49	65.0%	25.0%	0%	5% Type C Pitch 5% Coal Tar	22.2	75.6	
8/01/90	50	68.4%	26.3%	0%	5.3% Coal Tar	28.5	58.0	
8/02/90	51	90.0%	0.0%	0%	10% Resin	18.1	49.2	Note 4
8/06/90	54	65.0%	25.0%	0%	8% Liquid, 2% Powder	30.5	46.7	Note 5

Note 1 - Evaluated by U.E.C.
Note 2 - Calcined MGU char, 16 mesh.
Note 3 - Hot Briquetted, heated feedstock to 460°C for 15 minutes.
Note 4 - Char was 80% minus 16 mesh and 10% minus 60 mesh.
Note 5 - Char was 55% minus 30 mesh and 10% minus 60 mesh.

Table 3.41
Ultimate Analysis of Raw Coalite Liquid (as received)

	% by wt
Carbon	83.16
Hydrogen	8.52
Nitrogen	1.01
Oxygen	6.15
Sulphur	0.95
Water	1.57

Table 3.42
CTC Coal Liquids Comparison
(Chu et al., 1989)

I. Ultimate Analysis*, wt %		
	CTC Pro-Blend	Coalite
Carbon	85.01	84.49
Hydrogen	9.16	8.48
Oxygen	4.68	4.83
Nitrogen	0.67	1.03
Sulfur	0.50	0.97
H/C Atomic Ratio	1.284	1.196
Moisture	0.94	1.57
Molecular Wt	220	218
MJ/kg	38.9	-

II. Distillation Data (Dry Basis), Vol. %		
	CTC Pro-Blend	Coalite
Naphtha (IBP - 177°C)	4.7	1.4
Diesel (177° - 344°C)	60.1	49.1
Resid (B.P. > 344°C)	35.2	49.5

* Note: Data on dry basis, except moisture, molecular Wt, & MJ\kg, which are "as received".

Table 3.43

Commercial Petroleum Diesel and CTC Coal Diesel Fuels

(Miin et al., 1989)

	Commercial Diesel No. 2	CTC Diesel No. 2
Gravity (°API)	33.3	35
Flash Point (°C)	66	66
Cloud Point (°C)	- 10	- 9
Sulfur (wt %)	0.45	0.38
Ash (wt %)	< 0.01	< 0.01
Cetane No.	47	49
Kinematic Viscosity (40°C) cSt	3.1	2.8
ASTM Distillation	(See Figure 3.44)	

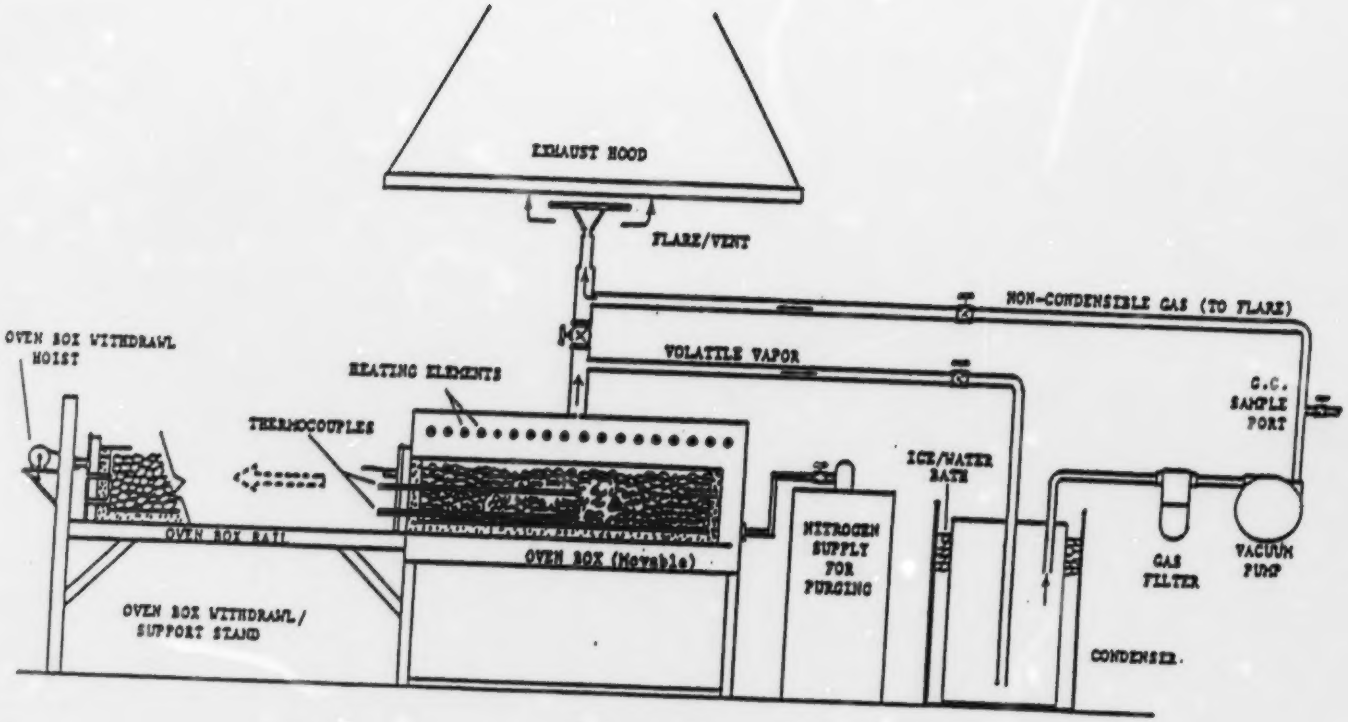

EXHAUST HOOD
FLARE/VENT
NON-CONDENSIBLE GAS (TO FLARE)
VOLATILE VAPOR
OVEN BOX WITHDRAWL HOIST
HEATING ELEMENTS
THERMOCOUPLES
G.C. SAMPLE PORT
ICE/WATER BATH
NITROGEN SUPPLY FOR PURGING
GAS FILTER
VACUUM PUMP
OVEN BOX RAIL
OVEN BOX (Movable)
OVEN BOX WITHDRAWL/ SUPPORT STAND
CONDENSER.

Figure 3.41

The CTC Char Briquette Facility

(Chu et al., 1989)

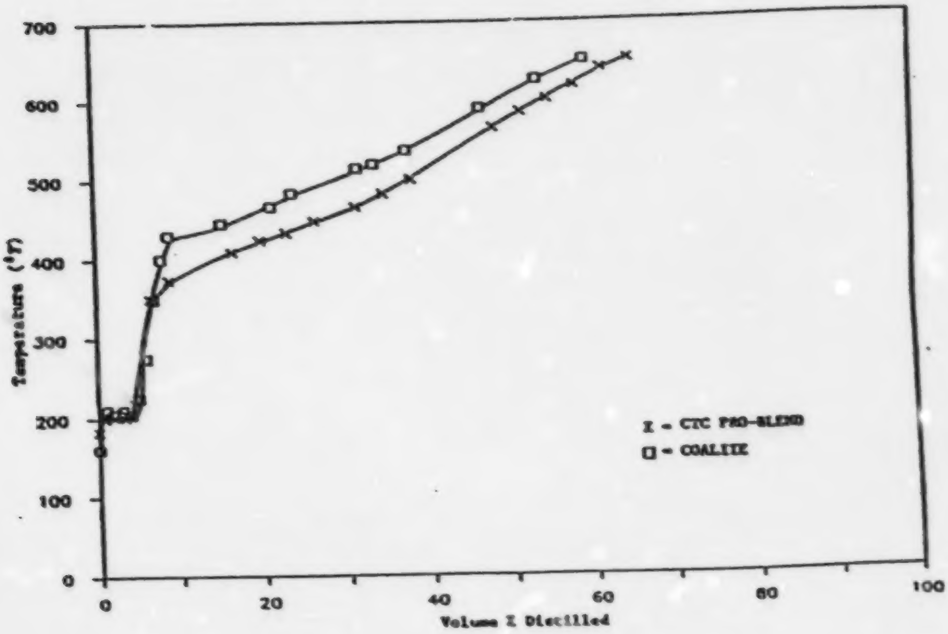


Figure 3.42

CTC Coal Derived Liquid Distillation Curve

(Chu et al., 1989)

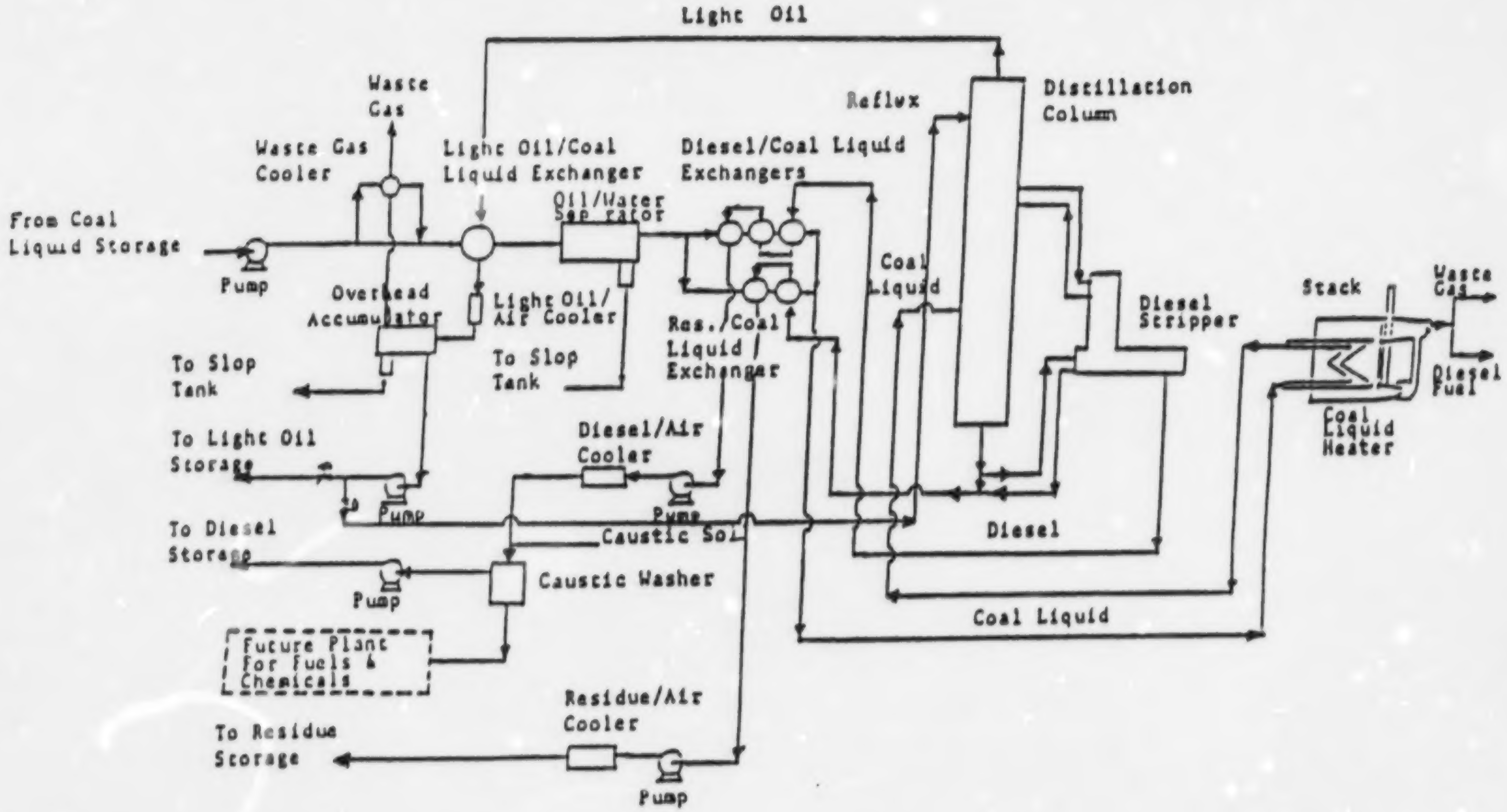

Light Oil
Waste Gas
Waste Gas Cooler
Light Oil/Coal Liquid Exchanger
Diesel/Coal Liquid Exchangers
Reflux
Distillation Column
From Coal Liquid Storage
Oil/Water Separator
Pump
Overhead Accumulator
Light Oil/Air Cooler
Coal Liquid
Res./Coal Liquid / Exchanger
Diesel Stripper
Stack
Waste Gas
Diesel Fuel
Coal Liquid Heater
To Slop Tank
To Slop Tank
To Light Oil Storage
Diesel/Air Cooler
Pump
Pump
To Diesel Storage
Caustic Sol
Diesel
Pump
Caustic Washer
Coal Liquid
Future Plant for Fuels & Chemicals
Residue/Air Cooler
To Residue Storage
Pump

Figure 3.43

The CTC Liquid Coal Refinery Processing Flowsheet

(Miln et al., 1989)

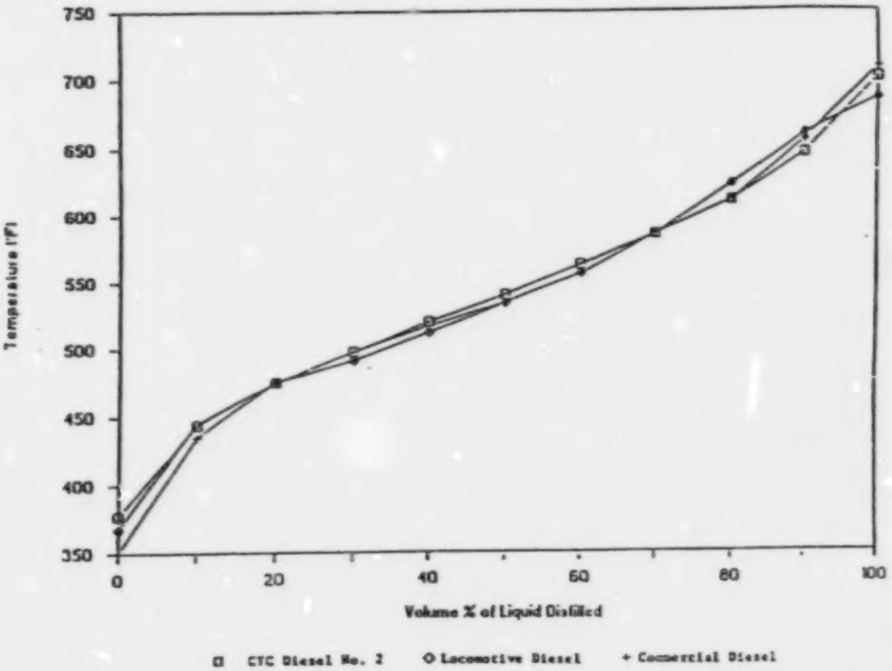


Figure 3.44

The CTC Diesel # 2 Distillation Curves

(Miin et al., 1989)

Tests have been carried out (Miin et al., 1989) on CTC-diesel No. 2 (Figure 3.44) in a number of engines including a 15 HP 3 cylinder lawn mower and 300 HP coal truck. The results of these tests were not discussed. The Southern Research Institute carried out 24 hours of tests using a 2 cylinder two stroke diesel locomotive engine. Cylinder pressure versus injection timing data for the CTC-diesel and a petroleum product showed very similar performance.

CTC has also produced a blend of coal-derived liquids and petroleum based diesel that they call CTC D-2. The analysis of this fuel is shown in Table 3.44.

CTC is marketing (Wright et al., 1990) a blend of coal derived liquids and other fuels at nine service stations in southwestern Virginia.

The 454 kg/h (coal process rate) process development unit (PDU) was reported to be approximately 90% completed by November, 1990 and nearly operational in January, 1991.

The PDU reactor consists of twin interfolded screw conveyors that are enclosed in a gas fired furnace (Wright et al., 1990). The screws are 40 cm OD fabricated on 25 cm OD hollow shafts that are 4.9 m long. The shafts have electric radiant heaters located inside.

The screws will turn at approximately 15 rpm and will intermittently reverse direction of rotation to remove caked coal from the surface and to increase the coal residence time. A 19 kW hydraulic drive will turn the screws. The gas fired and electric heating have independent controls. The reactor is divided into two separate heating zones and operated at slightly above atmospheric pressure.

The first zone of the reactor will see the coal heated to 205°C during which drying will take place. The temperature in the second zone is 454°C. The lump char from the reactor will be cooled in a water jacketed auger to 150°C. In the present system configuration, the char will be stored. The moisture and volatile hydrocarbons are recovered and separated in scrubber/condensers. The non-condensables will be flared in the PDU operation. These gases would be used to provide process heat in a commercial operation.

The operation of the PDU will begin in early 1991. Coal derived liquids will be generated and evaluated. Char from the PDU will be used to produce coke.

ENCOAL

The Encoal (Frederick et al., 1990) process uses a rotary grate reactor that is not suitable for caking coals. The products are a liquid with properties similar to a low sulphur # 6 fuel oil and a high heating value low sulphur solid fuel. Construction of a 1000 tonne/day plant is under way.

Table 3.44

Properties of the CTC D-2 Fuel

Property	1990 Procedure	1990 Results
Gravity, °API, 16°C	D1298	29.8
Specific Gravity, 16°C		.8767
Density, US Gallon, 16°C	7.275	7.303
Distillation, °C	D86	
IBP b/5		179/200
10/20		218/236
30/40		249/260
50/60		270/279
70/80		291/303
90/95		321/334
EP°		345
Recovery, %		-
Residue, %		-
Viscosity	D445	2.8
cSt @ 104°F (40°C)		
SUS @ 104°F (40°C)		
Flashpoint, °F	D93	154
Elemental Analysis, %:		
Carbon	CEC 240°	86.42
Hydrogen		12.01
Sulfur	XRF	0.506
Nitrogen	Chemiluminescent	0.136
Oxygen (difference)	Difference	0.92
Hydrocarbon Type , (FIA) %	D1319	
Aromatics		45.5
Olefins		1.2
Saturates		53.3

Table 3.44 (Continued)

Properties of the CTC D-2 Fuel

Property	1990 Procedure	1990 Results
Heat of Combustion	D240	
Gross, Btu/lb		19,023
Gross, MJ/kg		44.2
Net, Btu/lb		17,928
Net, MJ/kg		41.7
Accelerated Stability, mg/100 ml	D2274	6.2
Carbon Residue, 10% Btms, wt %	D524	0.35
Particulates	D2276	12.6
Metals, ppm in Fuel:	ICP	
Vanadium		< 1
Potassium		< 25
Nickel		< 1
Iron		< 1
Cetane Index	ASTM D976	40
Compatibility Rating, Procedure B with Cat H Diesel Fuel	Not Available	NA

Numbers preceded by a D are ASTM
IBP is the initial boiling point
EP is the end point
ICP is inductively coupled plasma

Institute of Gas Technology (IGT)

The process that is being developed at IGT is a cooperative venture between IGT, Peabody Holding Company (representative of the largest coal producer in the USA), Bechtel National Inc. and Caterpillar, Inc. The process is intended to produce co-products consisting of solids and liquid chemicals. The liquid chemicals may be used to produce electrode pitch binder or chemical feedstocks. The solids would be used to produce formed coke for metallurgical applications, smokeless fuel for displacing firewood and activated char for gas phase contaminant removal.

Four major tasks have been undertaken:

Task 1	Literature survey of mild gasification processes, co-products upgrading and utilization, and market assessment.
Task 2	Bench-scale mild gasification study
Task 3	Bench-scale char upgrading study
Task 4	System integration studies

Task 1, literature reviews and market surveys, identified technical process directions for commercialization of mild gasification. Potential markets for solid products in the US were identified as follows:

- metallurgical form coke at about 30 x 10^6 tonne/yr
- smokeless fuel at about 50 to 200 x 10^6 tonne/yr
- activated char at about 1 to 2 x 10^6 tonne/yr

In Task 2, the design of a 45 kg/h bench scale process research unit (PRU) began in April, 1988 with completion in February, 1989. Gasification testing began in May of 1989. An isometric view of the PRU is shown in Figure 3.45 (Onisschak et al., 1990). In Task 4 the full-stream condensate system shown in Figure 3.46 was added and operated.

IGT has called their bench scale gasification unit a process research unit (PRU). The PRU has a fluidized bed (lower) section that is 20 cm ID and 2.43 m high. An entrained bed sits on top of the FB that is 10 cm ID and 3.96 m high. Heating is provided from electrical heaters that surround the bed and electrical heaters used to reheat the fluidization gas. Coal feed can be at the base of the FB section or in the freeboard section. Solids discharge is either from the bottom of the bed or in an overflow port 61 cm above the distributor plate. The conceptual commercial reactor would obtain its process heat from the recycle of hot char and/or by high-temperature flue gases derived from burning a portion of the coal derived gases.

The following is a description of the PRU as it was operated in Task 2.

The product gas stream is maintained at 537°C using electrical heaters in the solids separation cyclones, which is above the tar condensation temperature.

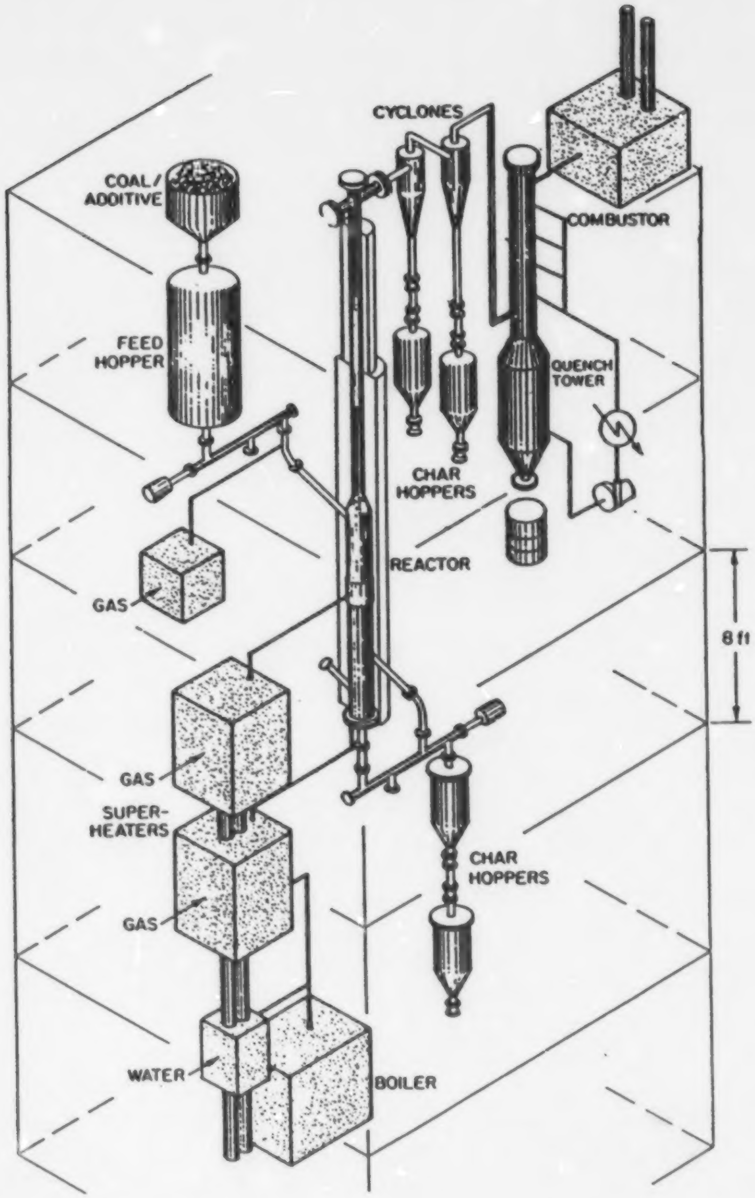

CYCLONES
COAL/ ADDITIVE
COMBUSTOR
FEED HOPPER
QUENCH TOWER
CHAR HOPPERS
REACTOR
GAS
8 ft
GAS
SUPER-HEATERS
CHAR HOPPERS
GAS
WATER
BOILER

Figure 3.45

The IGT Process Research Unit

(Knight et al., July 1990)

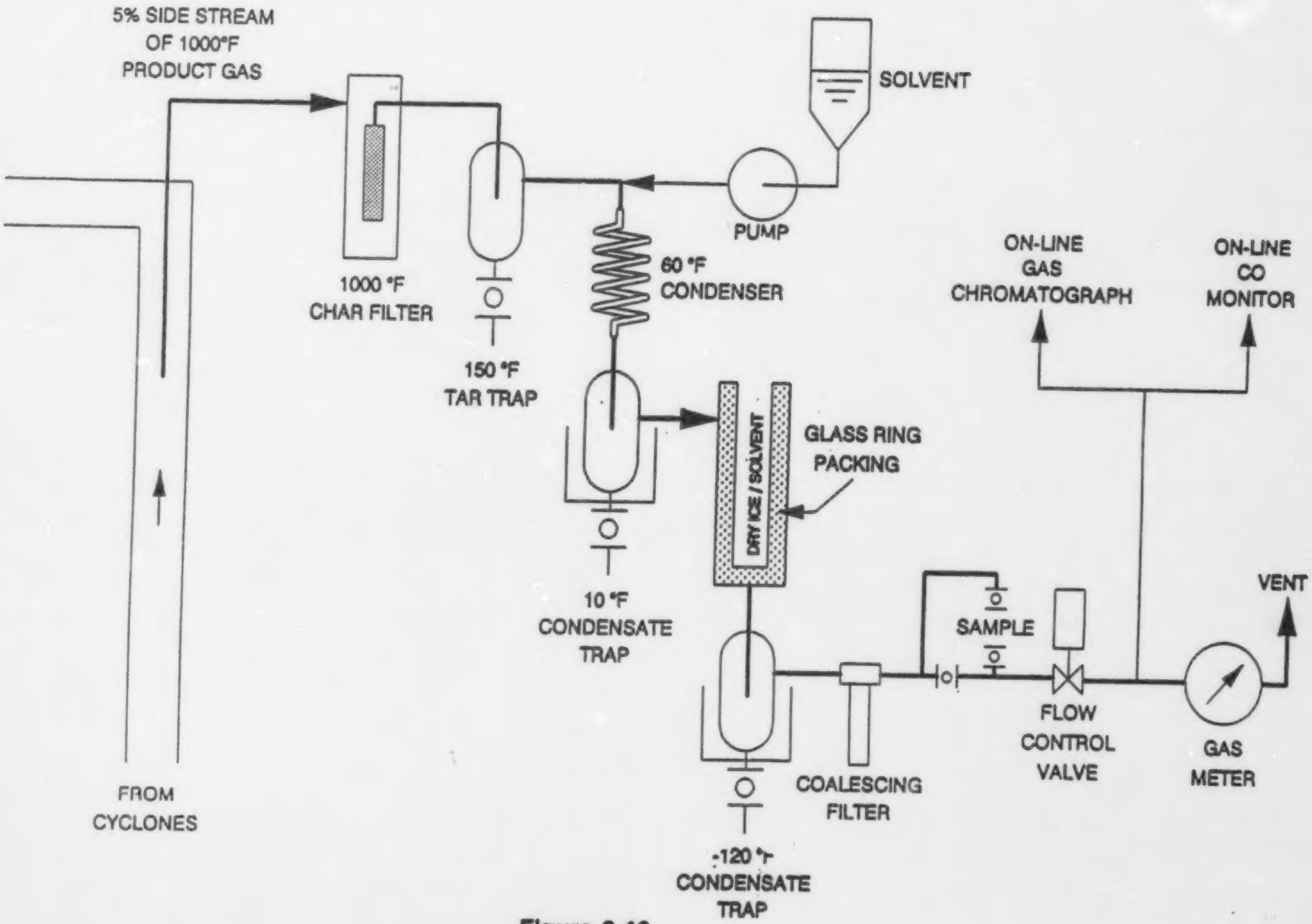

5% SIDE STREAM
OF 1000°F
PRODUCT GAS
FROM
CYCLONES
1000 °F
CHAR FILTER
150 °F
TAR TRAP
SOLVENT
PUMP
60 °F
CONDENSER
10 °F
CONDENSATE
TRAP
DRY ICE / SOLVENT
GLASS RING
PACKING
-120 °F
CONDENSATE
TRAP
COALESCING
FILTER
SAMPLE
FLOW
CONTROL
VALVE
GAS
METER
VENT
ON-LINE
GAS
CHROMATOGRAPH
ON-LINE
CO
MONITOR

Figure 3.46

The IGT Slipstream Sampling System

(Knight et al., July 1990

A slipstream of hot gas is sampled downstream of the cyclones and after cooling and filtering is analyzed using gas chromatography (Figure 3.46). A solvent (methylene chloride) quench system is used to collect tar aerosols.

Task 2 saw the testing of two coals, an Illinois No. 6 (Ill) and a metallurgical-grade blended coal from West Virginia (WV). Both coals were selected as representative of fines streams from preparation plants. Coke breeze was mixed with the coal to simulate the char recycle of a commercial operation. Analysis of the coals and the breeze is shown in Table 3.45. The Illinois coal had an FSI of 11/2 and the WV 51/2. The latter being more representative of the caking qualities of Nova Scotia coals.

Knight et al. (July 1990) reports that larger scale FB reactor systems would have larger inventories of char in the bed which would negate the need to blend the caking coal feed with char or breeze. Preliminary tests resulted in modifications being made to the fluidization grid as shown in Figure 3.47. The feed inlet port was moved closer to the fluidization grid to improve the fuel mixing as it enters the bed and prevent agglomeration of caking coal particles. The central fluidization jet was added to improve bed mixing.

The following analysis of the IGT PRU will be primarily focused on the results of the WV coal tests. Table 3.46 shows a summary of the Task 2 PRU gasification tests for the WV coal tests. The solids feed rates have been reduced from those used in the Illinois coal tests and ranged from 24 to 32 kg/h. The design input rate was approximately 45 kg/h. Subsequent data are reported in this report for tests MG-28 and MG-30. These tests had successful steady state periods that allowed the collection of significant data. Table 3.47 shows the analysis of the solid streams from the tests using the WV coal. The blocked feed line noted in test MG-27 was eliminated by reducing the temperature of the feed sweep gas from 366°C to 260°C. The feed block that occurred in test MG-29 was subsequently remedied by adding a mechanical push rod to break off agglomerates in the feed port. Some agglomeration of material was seen around thermocouples that protrude into the bed in test MG-30. A deeper bed (1.07 m) was used in test MG-31. Intermittent blockages were observed in the bed overflow nozzle. The material removed from the overflow was larger than the previous test (3.2 mm). The diluent shown in Table 3.47 is coke breeze. The second cyclone solids had a higher volatile and ash content than the first in both tests MG-28 and 30.

Subsequent evaluation of the feed hopper performance revealed that the addition of the coal/breeze mixture into the long feed hopper resulted in a segregation of the materials. This would affect the performance during the gasification of the caking coals and also affect the material balance data.

The solids analysis in Table 3.47, oils/tars elemental analysis in Table 3.48 and the gas components analysis in Table 3.49 were used to carry out the material balances for tests MG-28 and 30 as shown in Table 3.50. The coke breeze was considered to be non-reacting based on results of gasification tests carried out using the breeze only. The adjusted balances had the following procedure applied to allow a comparative analysis of the data (Knight et al., July 1990):

1. *The ash in the out[illegible]eams was balanced with the ash in the input feed by adjusting the weight of the fluidized-bed char collected during the test.*

Table 3.45

Analysis of the Coals and Coke Breeze Used in the IGT PRU Slipstream Task 2 Tests (Knight et al., July, 1990)

	Illinois No. 6[a]	West Virginia[b]	Coke Breeze	
			Batch 1[c]	Batch 2[d]
Proximate	wt % as fed to reactor			
Moisture	5.3	1.4	1.1	0.4
Volatile Matter	34.7	30.6	3.4	4.9
Ash	15.0	6.4	11.4	7.4
Fixed Carbon	45.0	61.6	84.1	87.3
Total	100.0	100.0	100.0	100.0
Ultimate	wt % dry basis			
Ash	15.8	6.4	11.6	7.4
Carbon	64.3	80.2	85.1	87.3
Hydrogen	4.2	5.0	0.4	1.2
Nitrogen	1.4	1.7	1.2	1.3
Sulphur	3.9	1.3	0.7	2.0
Oxygen (by diff)	10.4	5.4	1.0	0.8
Total	100.0	100.0	100.0	100.0
Sulphur Forms	wt % of total sulphur			
Pyritic	33.3	33.0	5.9	1.7
Sulphate	3.3	< 1.0	3.5	1.1
Sulphide	0.0	0.0	6.8	1.7
Organic	63.4	67.0	83.8	95.5
Total	100.0	100.0	100.0	100.0
HHV, MJ/kg (dry)	26.96	33.79	29.06	31.11
Free Swelling Index	1 ½	5 ½	0	0

a) Baldwin No. 1, Marissa and River King No. 6 Mines; proximate and ultimate analyses are averages of 4 samples; sulphur forms are from analysis of Test MG-16 feed coal

b) 55% No. 2 Gas Seam/45% Campbell's Creek Seam; proximate and ultimate analyses are averages of 3 samples; sulphur forms are "typical" values from supplier

c) Proximate and ultimate analyses, average of 3 samples; sulphur forms, 1 sample; used in tests up to and including Test MG-30

d) Proximate and ultimate analyses, average of 4 samples; sulphur forms, 1 sample; used in Test MG-31

Table 3.45 (Continued)

Analysis of the Coals and Coke Breeze Used in the IGT PRU Slipstream Task 2 Tests

(Knight et al., July, 1990)

	Illinois No. 6	West Virginia	Coke Breeze	
			Batch 1	Batch 2
Screen Size (μm)	wt % retained on screen			
1680	3.88	0.00	21.30	19.40
841	17.50	0.32	26.90	25.30
420	40.50	9.83	20.80	21.20
250	25.30	14.40	11.90	13.60
177	8.87	12.10	6.26	7.46
149	2.11	7.09	2.68	3.80
105	1.00	13.60	3.58	4.39
74	0.33	7.73	2.46	2.49
63	0.11	1.29	0.67	0.58
53	0.11	3.55	0.89	0.58
44	0.11	3.06	0.67	0.29
Pan	0.18	27.03	1.89	0.91
Total	100.00	100.00	100.00	100.00

Table 3.46

Summary of the IGT PRU Slipstream Gasification Tests from Task 2

(Knight et al., July, 1990)

	Test No.	
	MG-27	MG-28
Purpose of Test	First test with West Virginia Coal	Repeat MG-27 with feed sweep gas at lower temperature and higher flow
Feed Material	1:2 mixture coal[a]/coke	1:2 mixture coal[a]/coke
Feed Rate, kg/h	29.1	23.5
N_2 Rate, kg/h	72.9	78.7
Steam Rate, kg/h	3.7	3.7
Steam/Coal Ratio	0.38	0.47
Temperature, °C[b]	663	621
Pressure, kPag	69	69
Feed Location	Fluidized Bed	Fluidized Bed
Fluidized Bed Height, cm	67	43
Superficial Gas Velocity, m/s	1.1	1.1
Solids Residence Time, min.	-	15
Steady State Period, h	1.0	3.5
Remarks	Plug at end of feed nozzle to bed	Successful test operation

a) West Virginia coal
b) Average of four thermocouples

Table 3.46 (Continued)

Summary of the IGT PRU Slipstream Gasification Tests from Task 2

(Knight et al., July, 1990)

	Test No.		
	MG-29	MG-30	MG-31
Purpose of Test	Repeat MG-28 at lower temperature	Repeat MG-28 at lower temperature	Repeat MG-30 with 1:1 feed mixture
Feed Material	1:2 mixture coal[a]/coke	1:2 mixture coal[a]/coke	1:1 mixture coal[a]/coke
Feed Rate, kg/h	24.6	30.5	32.0
N_2 Rate, kg/h[b]	83.1	83.1	82.4
Steam Rate, kg/h	3.7	3.7	3.7
Steam/Coal Ratio	0.45	0.37	0.23
Temperature, °C[c]	532	566	564
Pressure, kPag	69	69	69
Feed Location	Fluidized Bed	Fluidized Bed	Fluidized Bed
Fluidized Bed Height, cm	73	73	114
Superficial Gas Velocity, m/s	1.1	1.1	1.1
Solids Residence Time, min.	-	17	19
Steady State Period, h	-	3.0	2.75
Remarks	Feed line plug at reactor stopped test	Successful test operation	Successful test but char overflow interruption

a) West Virginia coal
b) Not including purge nitrogen
c) Average of four thermocouples

Table 3.47

Analysis of the Solids Streams in the IGT PRU Slipstream Sample Test MG-28
(Test Temperature = 621°C)

(Knight et al., July, 1990)

	Feed Mixture	Coal	Diluent	Fluid Bed Char	1st Cyclone Char	2nd Cyclone Char	Carry Over Char
Proximate	wt %						
Moisture	0.80	0.16	1.12	0.17	0.19	0.02	1.58
Volatile Matter	9.23	20.81	3.44	3.94	7.92	17.37	13.46
Ash	11.55	11.77	11.44	12.77	16.10	31.56	8.31
Fixed Carbon	78.42	67.26	84.00	83.12	75.79	51.05	76.65
Total	100.00	100.00	100.00	100.00	100.00	100.00	100.00
Ultimate	dry wt %						
Carbon	83.02	78.82	85.12	83.51	77.05	57.54	91.56
Hydrogen	1.49	3.77	0.35	0.64	1.83	2.24	ND
Sulfur	0.98	1.54	0.70	0.95	1.52	0.18	ND
Nitrogen	1.29	1.45	1.21	1.36	1.34	1.40	ND
Oxygen (by diff)	1.58	2.64	1.05	0.75	2.13	7.07	ND
Ash	11.64	11.78	11.57	12.79	16.13	31.57	8.44
Total	100.00	100.00	100.00	100.00	100.00	100.00	100.00
Heating Value, MJ/kg	ND	33.79	29.06	ND	ND	ND	ND
Particle Density, g/cm^3	ND	ND	1.82	ND	ND	ND	ND
Particle Size Distribution							
(μm)	wt % retained on screen						
3360	ND	0.00	0.00	ND	ND	ND	ND
1680	-	0.00	19.30	-	-	-	-
841	-	0.32	31.40	-	-	-	-
595	-	-	-	-	-	-	-
420	-	9.83	37.00	-	-	-	-
250	-	14.40	11.90	-	-	-	-
210	-	-	-	-	-	-	-
177	-	12.10	0.17	-	-	-	-
149	-	7.09	0.00	-	-	-	-
125	-	-	-	-	-	-	-
105	-	13.60	0.00	-	-	-	-
88	-	-	-	-	-	-	-
74	-	7.73	0.00	-	-	-	-
63	-	1.29	0.17	-	-	-	-
53	-	3.55	0.00	-	-	-	-
44	-	3.06	0.00	-	-	-	-
Pan	-	27.03	0.16	-	-	-	-
Total	-	100.00	100.00	-	-	-	-

ND = analysis not done

Table 3.47 (Continued)

Analysis of the Solids Streams in the IGT PRU Slipstream Sample Test MG-30 (Test Temperature = 566°C)

(Knight et al., July, 1990)

	Feed Mixture	Coal	Diluent	Fluid Bed Char	1st Cyclone Char	2nd Cyclone Char	Carry Over Char
Proximate	wt %						
Moisture	0.88	0.40	1.12	0.23	0.53	0.59	0.59
Volatile Matter	9.88	22.76	3.44	4.55	9.04	21.84	19.77
Ash	11.23	9.85	11.44	12.21	14.70	43.20	4.62
Fixed Carbon	78.01	66.99	84.00	83.01	75.73	34.37	75.02
Total	100.00	100.00	100.00	100.00	100.00	100.00	100.00
Ultimate	dry wt %						
Carbon	82.15	76.21	85.12	82.71	77.40	44.02	95.47
Hydrogen	1.51	3.83	0.35	0.82	1.78	2.23	ND
Sulfur	0.96	1.48	0.70	0.94	1.40	1.01	ND
Nitrogen	1.34	1.60	1.21	1.39	1.35	0.98	ND
Oxygen (by diff)	3.03	6.99	1.05	2.17	3.49	10.79	ND
Ash	11.01	9.89	11.57	11.97	14.58	40.97	4.53
Total	100.00	100.00	100.00	100.00	100.00	100.00	100.00
Heating Value, MJ/kg	ND	33.79	29.06	ND	ND	ND	ND
Particle Density, g/cm^3	ND	ND	1.82	ND	ND	ND	ND
Particle Size Distribution (μm)	wt % retained on screen						
3360	ND	0.00	0.00	ND	ND	ND	ND
1680	-	0.00	19.30	-	-	-	-
841	-	0.32	31.40	-	-	-	-
595	-	-	-	-	-	-	-
420	-	9.83	37.00	-	-	-	-
250	-	14.40	11.90	-	-	-	-
210	-	-	-	-	-	-	-
177	-	12.10	0.17	-	-	-	-
149	-	7.09	0.00	-	-	-	-
125	-	-	-	-	-	-	-
105	-	13.60	0.00	-	-	-	-
88	-	-	-	-	-	-	-
74	-	7.73	0.00	-	-	-	-
63	-	1.29	0.17	-	-	-	-
53	-	3.55	0.00	-	-	-	-
44	-	3.06	0.00	-	-	-	-
Pan	-	27.03	0.16	-	-	-	-
Total	-	100.00	100.00	-	-	-	-

ND = analysis not done

Table 3.48

Oils/Tars Elemental Analysis and Simulated Distillation for the IGT PRU Slipstream Tests

(Knight et al., July, 1990)

Test Number	MG-28	MG-30
Test Temperature, °C	621	566
Elemental Analysis of Oils and Tars[a]	dry wt %	
Ash	1.14	0.77
Carbon	79.60	82.04
Hydrogen	7.09	7.34
Nitrogen	0.86	0.82
Sulphur	0.83	0.74
Oxygen (by diff)	10.48	8.29
Total	100.00	100.00
H/C Atomic Ratio	1.06	1.07
Simulated Distillation by Gas Chromatography[b]		
Cumulative wt % Recovered	Boiling Point, °C	
5	191	156
10	241	220
15	283	269
20	324	312
30	401	388
40	484	461
50	576	535
60	-	-
70	-	-
EP (end point)[c]	576	576
% Residue at EP	51.8	45.0

a) Determined by evaporation at 38°C, 15 - 20 mm Hg; light oils boiling below approximately 149°C are not included

b) Correction applied for heteroatom content of coal liquids, which is not accounted for in standard simulated distillation method for petroleum-based liquids

c) Characteristic of chromatographic column and method, not necessarily true end point of distillation

Table 3.49

Analysis of the Gas Components for the IGT PRU Slipstream Tests

(Knight et al., July, 1990)

Test Number	MG-28	MG-30
Test Temperature, °C	1149	1050
Component	Mol % in gas, nitrogen-free	
H_2	27.1	18.9
CO	3.8	1.9
CO_2	8.2	8.9
CH_4	45.2	51.3
C_2H_4	5.3	5.9
C_2H_6	5.4	6.6
C_3H_6	2.6	2.9
C_3H_8	1.0	2.0
H_2S	1.4	1.6
Total	100.0	100.0
Molecular Weight	17.6	19.3
High Heating Value, MJ/m^3	30.9	34.3

Table 3.50

Material Balances for the IGT Slipstream Test MG-28 (621°C)

(Knight et al., July, 1990)

	Unadjusted						
	C	H	O	N	S	Ash	Total
Input							
Feed Mixture	42.74	0.77	0.82	0.66	0.51	5.99	51.49
Steam + Moisture	-	0.97	7.66	-	-	-	8.63
Total	42.74	1.74	8.48	0.66	0.51	5.99	60.12
Output							
Char Mixture	42.59	0.32	0.40	0.69	0.48	6.52	51.00
Product Gas	0.82	0.24	0.25	0.00	0.03	-	1.34
Oils/Tars	2.89	0.26	0.36	0.03	0.03	0.00	3.57
Aqueous Condensate	0.00	0.37	2.92	0.00	0.00	0.00	3.29
Total	46.30	1.19	3.93	0.72	0.54	6.52	59.20
Output/Input	1.08	0.68	0.46	1.09	1.06	1.09	0.98

	Adjusted						
	C	H	O	N	S	Ash	Total
Input							
Feed Mixture	42.74	0.77	0.82	0.66	0.51	5.99	51.49
Output							
Char Mixture	39.10	0.30	0.35	0.64	0.44	5.99	46.82
Product Gas	0.82	0.24	0.25	0.00	0.03	-	1.34
Oils/Tars	2.82	0.25	0.35	0.03	0.03	0.00	3.48
Aqueous Condensate	0.00	- 0.02	- 0.12	- 0.01	0.00	-	- 0.15
Total	42.74	0.77	0.83	0.66	0.50	5.99	51.49
Output/Input	1.00	1.00	1.02	1.00	1.00	1.00	1.00

Diluent-Free Co-Product Yields, wt % Coal		
	dry	maf
Char	73.0	69.4
Gas	7.8	8.8
Oils/Tars	20.1	22.8
Water	- 0.9	- 1.0
Total	100.00	100.00

Table 3.50 (Continued)

Material Balances for the IGT Slipstream Test MG-30 (566°C)

(Knight et al., July, 1990)

	Unadjusted						
	C	H	O	N	S	Ash	Total
Input							
Feed Mixture	54.72	1.00	2.02	0.90	0.64	7.34	66.62
Steam + Moisture	-	0.98	7.81	-	-	-	8.79
Total	54.72	1.98	9.83	0.90	0.64	7.34	75.41
Output							
Char Mixture	47.30	0.47	1.24	0.79	0.55	6.85	57.20
Product Gas	0.73	0.21	0.19	0.00	0.03	-	1.16
Oils/Tars	3.58	0.32	0.35	0.03	0.03	0.00	4.31
Aqueous Condensate	0.00	0.84	6.68	0.01	0.00	0.00	7.53
Total	51.61	1.84	8.46	0.83	0.61	6.85	70.20
Output/input	0.94	0.93	0.86	0.92	0.95	0.93	0.93

	Adjusted						
	C	H	O	N	S	Ash	Total
Input							
Feed Mixture	54.72	1.01	2.02	0.89	0.64	7.34	66.62
Output							
Char Mixture	50.66	0.50	1.33	0.85	0.58	7.34	61.26
Product Gas	0.73	0.21	0.19	0.00	0.03	-	1.16
Oils/Tars	3.58	0.32	0.35	0.03	0.03	0.00	4.31
Aqueous Condensate	0.00	- 0.01	- 0.11	0.01	0.00	-	- 0.11
Total	54.97	1.02	1.76	0.89	0.64	7.34	66.62
Output/Input	1.00	1.02	0.87	1.00	1.00	1.00	1.00

Diluent-Free Co-Product Yields, wt % Coal		
	dry	maf
Char	76.0	73.4
Gas	5.2	5.7
Oils/Tars	19.3	21.4
Water	- 0.5	- 0.5
Total	100.00	100.00

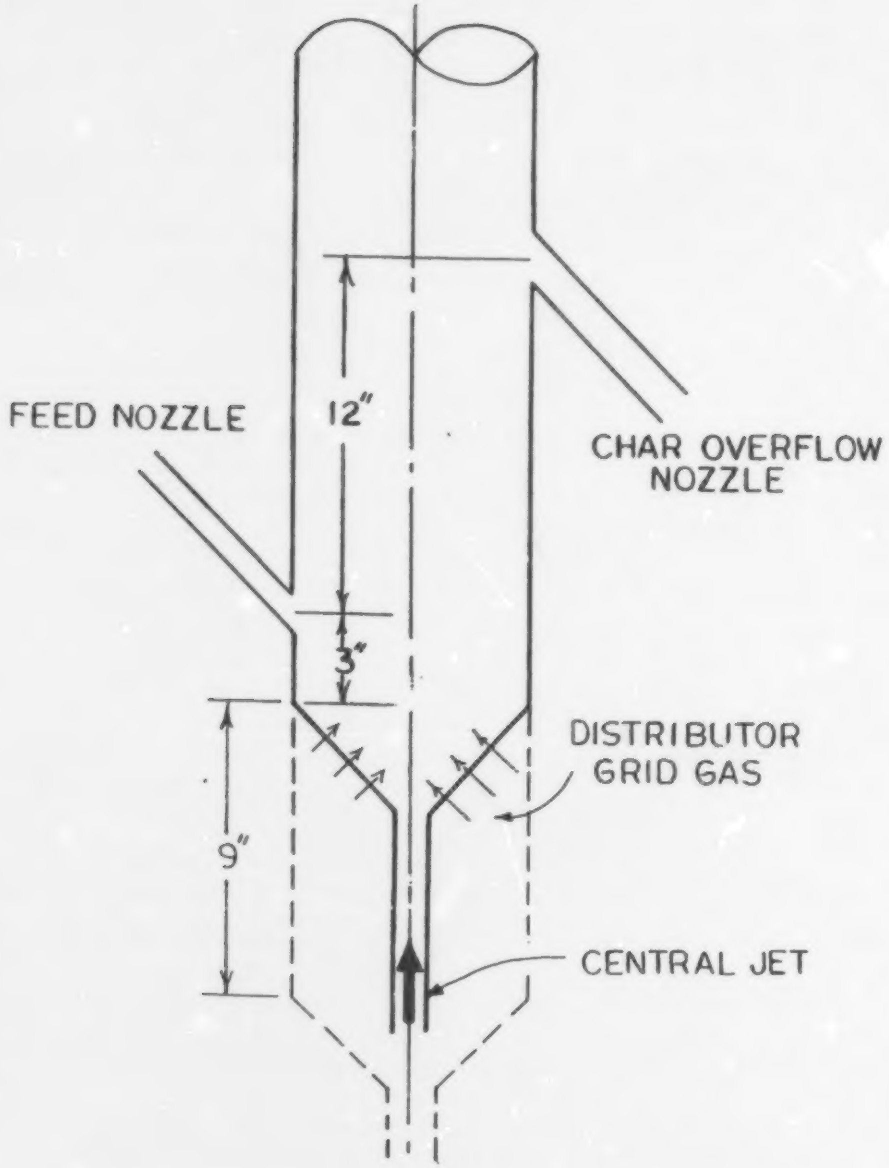

FEED NOZZLE
12"
CHAR OVERFLOW NOZZLE
3"
9"
DISTRIBUTOR GRID GAS
CENTRAL JET

Figure 3.47

The IGT PRU Fluidization Grid Modifications

(Knight et al., July 1990)

2. *The carbon in the output streams was balanced by adjusting the amount of oils and tars collected, assuming that the heavy components coating pipes in the slipstream sampling system cannot be quantitatively collected, and also that the sintered metal filter used to separate char from the gas stream provides an active surface that promotes thermal cracking of a portion of the heavier tars, resulting in deposition of tar-based carbon in the filter pores. In cases where the carbon balance after the first step was in excess of 100%, no adjustment was made.*

3. *Sulphur gases react reversibly with stainless steel system components, thus introducing a random error in the PRU sulphur balances, unaccounted sulphur was balanced by assuming its presence as* H_2S.

4. *Unaccounted nitrogen from the coal was assumed to form ammonia in the condensate stream. An analysis was performed for ammonia, but the volatility of that species at the low pH of the aqueous condensate makes it probable that a significant portion is lost during sampling and sample transport and storage.*

5. *The net chemical water make was determined by difference from a 100% total material balance. In some tests, the net water make is a negative number. It is unlikely that steam is actually consumed in the test, and the negative value for net water is probably the result of an accumulation of experimental error.*

The consideration made in step 3 above points to a need to change the material in a scaled up operation to prevent excessive corrosion.

Re-examination of the conditions for these tests show that the operational conditions that varied significantly were the reactor temperature (621 and 566°C respectively), the feed rate (23.5 and 30.5 kg/h respectively), the steam to coal ratio (0.47 and 0.37 respectively) and the fluidized bed depth (43 and 73 cm respectively). Table 3.47 indicates that there were no significant variations in the composition of the feed mixture. Variations in the actual blending of the feed material as fed to the bed may have an unknown affect on the balance data. Figure 3.48 shows the yields of char, oils/tars, gas and water for all of the IGT PRU tests.

An increase in the reactor temperature resulted in a slight increase in the oil/tars yield (22.4 and 22.8%). The higher yield oil/tars had a higher ash and oxygen content (Table 3.48). The H/C atomic ratio remained virtually unchanged at approximately 1.06. Table 3.51 shows the component analysis of the oil/tars. The data in this table more accurately represent the full range of condensable hydrocarbons from the process (refer to note a in Table 3.51).

The increased reactor temperature resulted in an increase in the gas yields (5.7 to 8.8%) as would be expected. Commercial operations using a coal-gas combustion product would affect this result somewhat. In so much that the gas is to be used to generate process heat, this would be of little consequence.

Table 3.51

Component Analysis of Full-Range Oils/Tars for IGT PRU Task 4 Tests With West Virginia Coal

(Knight et al., July, 1990)

Test Number	MG-28	MG-30
Test Temperature, °C	621	566
Component	wt % of Total oils/tars[a]	
Benzene	0.5	0.3
Toluene	0.7	0.6
Xylenes	0.5	0.6
Ethylbenzene	0.1	0.1
Indene	0.2	0.1
Styrene	0.05	0.1
Other Light Oils	7.3	8.7
Total Light Oil[b]	9.3	10.5
Phenol	0.4	0.4
Cresols	1.0	1.0
Xylenols	1.6	1.3
Naphthalene	0.6	0.1
Other Middle Oils	9.7	9.7
Total Middle Oil[c]	13.3	12.5
Heavy Oil[d]	10.8	11.2
Pitch[e]	66.6	65.8
Total Oils and Tars	100.0	100.0

a) Includes light oils that are not included in the oils and tars of Table 3.48
b) Atmospheric boiling point < 182°C; estimated from simulated distillation data
c) Atmospheric boiling point 182 to 310°C
d) Atmospheric boiling point 310 to 399°C
e) Atmospheric boiling point > 399°C

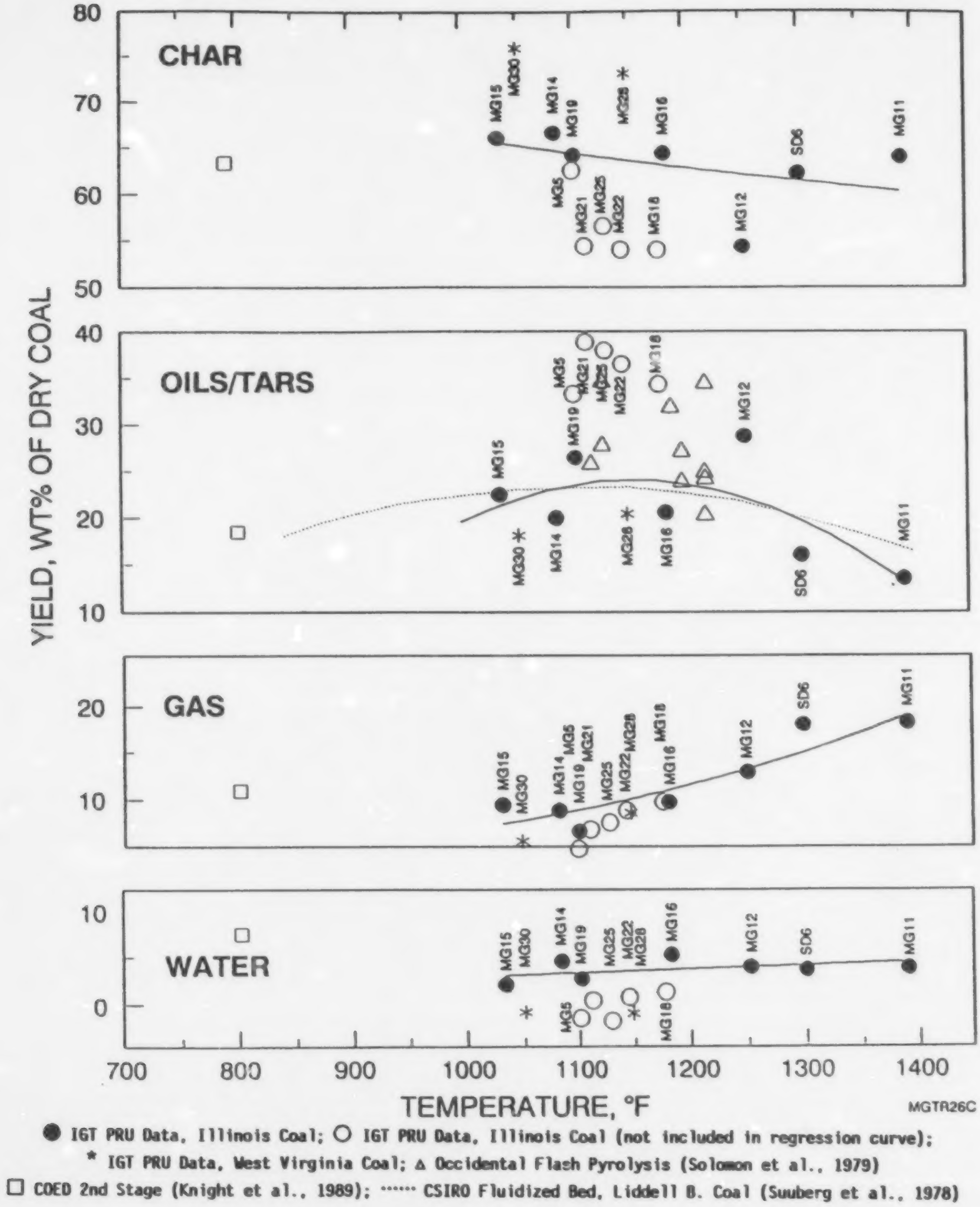


Figure 3.48

IGT, PRU Slipstream, Effect of Temperature on the Yields

(Knight et al., July 1990)

Babu et al. (1990) reported that form coke was produced from a 1:1 blend of hot char and coal. Knight et al., December 1990, notes that the char from test IST-10 was used to make form coke. It is not clear if this is the same form coke that Babu describes. Pitch was not used as a binder in order that its higher value might be obtained in the pitch form. The tensile strength of the formed coke was within the range seen in presently used commercial cokes (2.3 MPa). The Bethlehem Steel Company test procedure to measure and compare CO_2 reactivity indicated that the form coke was comparable to commercial products.

Smokeless fuel (for domestic use) was prepared as briquettes from char with approximately 12% pitch. Limestone or dolomite was added to act as a sulphur sorbent at Ca/S molar ratio of 2:1. In this form, 84 to 88% of the sulphur in the fuel is retained in the solid form after combustion.

Activated chars were prepared by steam treating the char at 850°C. The activated chars had surface areas and adsorption performances that were similar to commercial active carbons. The pore size in the activated chars was smaller than commercial products indicating that they might adsorb low molecular weight substances efficiently.

Task 4 of the IGT PRU investigation evaluated the integration of a full stream condensate system as shown in Figure 3.49 (Knight et al., December 1990). The collection system is operated such that the low boiling fraction of the condensable liquids are not collected but pass into an incinerator to avoid discharge to the atmosphere. The yield of the low boiling point fraction can be determined using the slipstream apparatus. Initial operation of the full stream condensate system resulted in the following modifications:

- a mesh demister was added to the top of the tower to enhance the removal of tar aerosols
- three of the original spray nozzles were used
- water as a quench liquid was replaced with xylene, which was considered safe to use and could be separated from the products
- the pressure let down valve at the exit of the tower was moved closer to the tower to allow it to operate at a higher temperature, this resulted in the need to increase the exit piping diameter due to its operation at nearly atmospheric pressure as opposed to the reactor pressure
- a refrigeration chiller was added to the xylene heat exchanger

The following is a presentation and discussion of the test results using the WV coal after the modifications noted above had been made to the condensate system. Table 3.52 shows the operation conditions using the WV coal and the full-stream condensate system. One of the objectives of the tests was to use char from previous tests and reduce the coke content in the recycle char diluent.

Table 3.52

Summary of the IGT PRU Gasification Tests Using the Full-Stream Condensate System, Task 4, West Virginia Coal

(Knight et al., December, 1990)

	Test No.		
	IST-10	IST-11	IST-12
Purpose of Test	Conduct WV coal test, reduce coke by using MG-31 char diluent	Reduce coke content further using IST-10 char diluent	Conduct 12-hour steady-state test
Feed Material	50% coal[a]/33% char/ 17% coke	1:1 mixture coal[a]/char	1:1 mixture coal[a]/coke
Feed Rate, kg/h	24.3	33.6	28.6
Steam Rate, kg/h	0.0	0.0	0.0
Temperature, °C	597	599	593
Pressure, kPag	67	69	69
Feed Location	Fluidized Bed	Fluidized Bed	Fluidized Bed
Fluidized Bed Height, cm	58	69	49
Superficial Gas Velocity, m/s	1.4	1.3	1.3
Solids Residence Time, min	13	9	10
Steady-State Period, h	4.5	3.5	12
Remarks	Successful test operation	Successful test operation	Successful test operation

a) West Virigina Coal
b) Illinois No. 6 coal

Table 3.52 (Continued)

Summary of the IGT PRU Gasification Tests Using the Full-Stream Condensate System, Task 4, West Virigina Coal

(Knight et al., December, 1990)

	Test No.	
	IST-13	IST-14
Purpose of Test	Reduce Coke in feed by using IST-12 char diluent	Obtain higher temperature data, feed using IST-12
Feed Material	1:1 mixture coal[a]/char	1:1 mixture coal[a]/char
Feed Rate, kg/h	27.8	27.5
Steam Rate, kg/h	0.0	0.0
Temperature, °C	584	691
Pressure, kPag	69	82
Feed Location	Fluidized Bed	Fluidized Bed
Fluidized Bed Height, cm	70	49
Superficial Gas Velocity, m/s	1.4	1.4
Solids Residence time, min	11	9
Steady-State Period, h	5.0	7.0
Remarks	Successful test operation	Successful test operation

a) West Virginia Coal

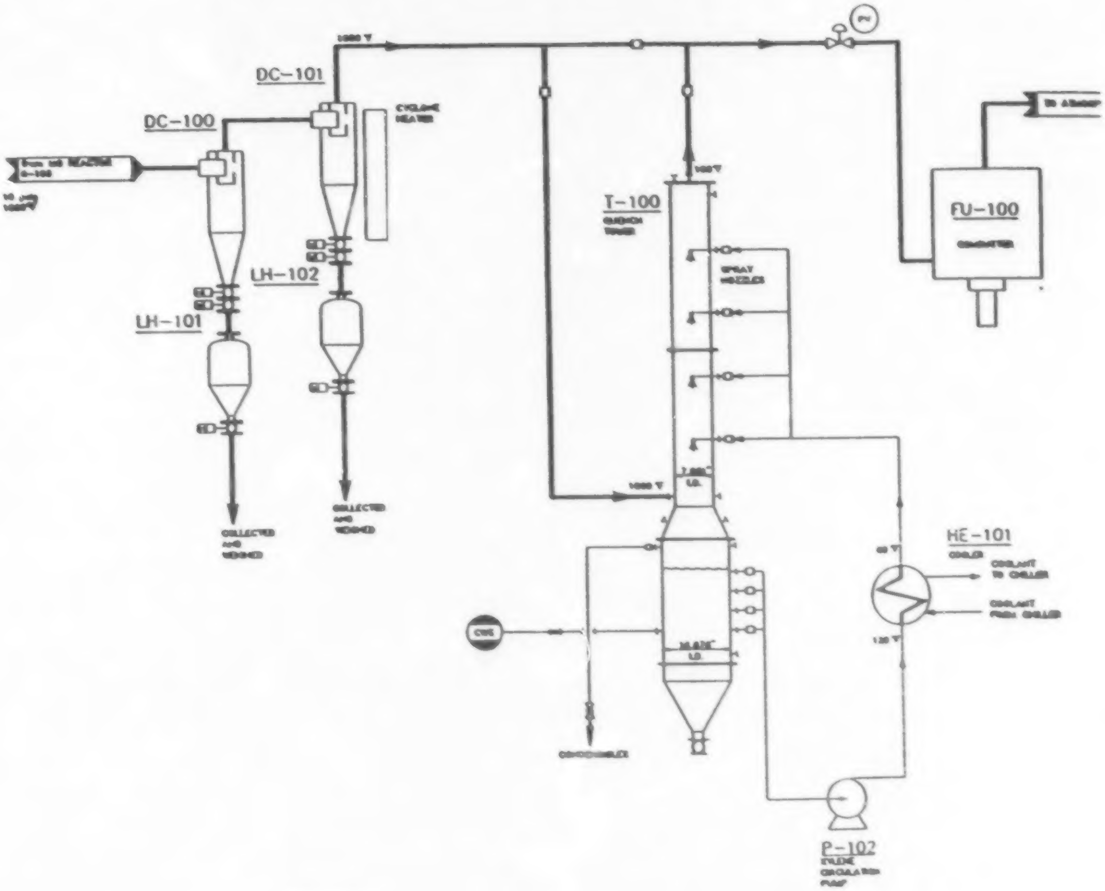

DC-101
CYCLONE
HEATER
DC-100
LH-102
LH-101
COLLECTED
AND
WEIGHED
COLLECTED
AND
WEIGHED
T-100
SPRAY
NOZZLES
FU-100
HE-101
COOLANT
TO CHILLER
COOLANT
FROM CHILLER
P-102

Figure 3.49

The IGT PRU Full-Stream Condensate System

(Knight et al., December 1990)

A simplified mass balance procedure was used in analyzing these data. The oils/tars yield was adjusted to force the overall mass balance to 100% on a moisture ash free basis. Tests were conducted with diluent only to allow a correction for volatiles released from the diluent in the yield analysis. The mass balance data for these tests are shown in Tables 3.53 to 3.57. **It must be emphasized that the oils/tars data used in the mass balances were derived from the slipstream data and not the full-stream condensate system.** Comparisons made between the Task 2 and the Task 4 results must consider the following changes in the Task 4 operation conditions:

- no steam was used in the fluidizing gas
- solids residence times were lower
- coal derived char was used to replace some of the coke breeze used as diluent

Figure 3.50 shows a comparison of the co-product yield data from Task 2 and 4. The Task 2 data has been re-evaluated using the same mass balance procedures and the data from the two sets of tests is comparable.

Table 3.58 shows the oils/tars elemental analysis and simulated distillation data for the IGT PRU Task 4 tests with WV coal. The simulated distillation data indicates that a higher boiling point oil/tar with more residue is produced as the coal derived char component of the diluent increases (with bed temperature held constant). The lack of any change in the aromaticity in the oils/tars indicates the increase in the boiling point range is due to changes in the product after it has been collected from the system. IGT has speculated that the high heteroatom content, especially oxygen, which results in a high sensitivity to aging, may have resulted in changes in the liquid properties during storage.

Table 3.59 shows the component analyses of the full-range oils/tars indicating that the majority of the oils/tars are in the pitch range (i.e. atmospheric boiling > 400°C).

Table 3.60 shows the gas composition for the IGT PRU Task 4 tests using WV coal. These data are averages of both on-line and off-line gas chromatography and are reported on a N_2 free basis.

The analysis of the full-stream oils/tars shown in Table 3.61 have been carried out on the product prepared in the following manner. The oils/tars collected in the xylene quench were distilled to remove the xylene. The tar laden quench liquid was filtered and washed with fresh xylene. The filtrate was then distilled to separate the xylene. The undistilled filtrate, overheads and the pot residue were analyzed using GC; the residue was also subjected to elemental analysis. The tar residue component contains less than half of the original components that boil between 203 and 227°C due to the xylene distillation step temperature. This loss would include a significant portion of the monoaromatic phenols. Tests carried out using other coals with similar rank and elemental composition yielded condensable products that were much less viscous. *This indicates the need for testing of the coal to be considered for evaluation in order that the condensable recovery system be properly designed.*

Table 3.53

Material Balances for the IGT Full-Stream Test IST-10 (621°C)
Coal: West Virginia, Fluidization Gas: 100% N_2, Temperature: 598°C

(Knight et al., December, 1990)

	Unadjusted						
	C	H	O	N	S	Ash	Total
Input							
Feed	21.20	1.11	1.15	0.39	0.33	2.01	26.19
Diluent	22.45	0.43	0.64	0.40	0.44	2.34	26.70
Steam + Moisture	-	0.06	0.54	-	-	-	0.6
Total	43.65	1.60	2.33	0.79	0.77	4.35	53.49
Output							
Char + Diluent	35.18	0.69	1.04	0.59	0.68	4.19	42.37
Product Gas	1.11	0.31	0.21	0.00	0.04	-	1.67
Oils/Tars	5.04	0.47	0.63	0.04	0.05	0.02	6.25
Aqueous Condensate	0.00	0.15	1.29	0.00	0.00	-	1.44
Total	41.33	1.62	3.17	0.63	0.77	4.21	51.73
Output/Input	0.95	1.01	1.36	0.80	1.00	0.97	0.97

	Adjusted						
	C	H	O	N	S	Ash	Total
Input							
Feed	21.20	1.11	1.15	0.39	0.33	2.01	26.19
Diluent	22.45	0.43	0.64	0.40	0.44	2.34	26.70
Steam + Moisture	-	0.06	0.54	-	-	-	0.60
Total	43.65	1.60	2.33	0.79	0.77	4.35	53.49
Output							
Char + Diluent	35.18	0.69	1.04	0.59	0.68	4.19	42.37
Product Gas	1.11	0.31	0.21	0.00	0.04	-	1.67
Oils/Tars	6.21	0.58	0.78	0.05	0.06	0.02	7.70
Aqueous Condensate	0.00	0.15	1.29	0.00	0.00	-	1.44
Total	42.50	1.73	3.32	0.64	0.78	4.21	53.18
Output/Input	0.97	1.08	1.42	0.81	1.01	0.97	0.99

Diluent-Free Co-Product Yields, wt % Coal		
	dry	maf
Char	60.4	57.8
Gas	6.4	6.9
Oils/Tars	29.4	31.8
Water	3.2	3.5
Total	99.4	100.0

Table 3.54

Material Balances for the IGT Full-Stream Test IST-11 Coal: West Virginia, Fluidization Gas: 100% N_2, Temperature: 599°C

(Knight et al., December, 1990)

	Unadjusted						
	C	H	O	N	S	Ash	Total
Input							
Feed	28.06	1.59	1.37	0.56	0.30	2.55	34.43
Diluent	28.99	0.58	0.86	0.49	0.56	3.45	34.93
Steam + Moisture	-	0.11	0.84	-	-	-	0.95
Total	57.05	2.28	3.07	1.05	0.86	6.00	70.31
Output							
Char + Diluent	47.36	1.11	1.44	0.93	0.670.	6.05	57.56
Product Gas	1.18	0.33	0.32	0.00	03	-	1.86
Oils/Tars	5.28	0.47	0.68	0.05	0.03	0.00	6.51
Aqueous Condensate	0.00	0.21	1.60	0.00	0.00	-	1.81
Total	53.82	2.12	4.04	0.98	0.73	6.05	67.74
Output/Input	0.94	0.93	1.32	0.93	0.85	1.01	0.96

	Adjusted						
	C	H	O	N	S	Ash	Total
Input							
Feed	28.06	1.59	1.37	0.56	0.30	2.55	34.43
Diluent	28.99	0.58	0.86	0.49	0.56	3.45	34.93
Steam + Moisture	-	0.11	0.84	-	-	-	0.95
Total	57.05	2.28	3.07	1.05	0.86	6.00	70.31
Output							
Char + Diluent	47.36	1.11	1.44	0.93	0.67	6.05	57.56
Product Gas	1.18	0.33	0.32	0.00	0.03	-	1.86
Oils/Tars	7.27	0.65	0.96	0.08	0.05	0.01	9.02
Aqueous Condensate	0.00	0.21	1.60	0.00	0.00	-	1.81
Total	55.81	2.30	4.32	1.01	0.75	6.06	70.25
Output/Input	0.98	1.01	1.41	0.96	0.87	1.01	1.00

Diluent-Free Co-Product Yields, wt % Coal		
	dry	maf
Char	66.1	63.2
Gas	5.4	5.8
Oils/Tars	26.2	28.3
Water	2.5	2.7
Total	100.2	100.0

Table 3.55

Material Balances for the IGT Full-Stream Test IST-12
Coal: West Virginia, Fluidization Gas: 100% N_2, Temperature: 594°C

(Knight et al., December, 1990)

	Unadjusted						
	C	H	O	N	S	Ash	Total
Input							
Feed	23.34	1.49	1.51	0.46	0.30	1.73	28.83
Diluent	24.98	0.09	0.12	0.31	0.16	3.48	29.14
Steam + Moisture	-	0.06	0.47	-	-	-	0.53
Total	48.32	1.64	2.10	0.77	0.46	5.21	58.50
Output							
Char + Diluent	39.97	0.49	0.65	0.64	0.34	5.68	47.77
Product Gas	1.12	0.34	0.08	0.00	0.01	-	1.55
Oils/Tars	7.47	0.70	0.92	0.08	0.07	0.07	9.31
Aqueous Condensate	0.00	0.28	2.17	0.00	0.00	-	2.45
Total	48.56	1.81	3.82	0.72	0.42	5.75	61.08
Output/Input	1.00	1.10	1.82	0.94	0.91	1.10	1.04
	Adjusted						
	C	H	O	N	S	Ash	Total
Input							
Feed	23.34	1.49	1.51	0.46	0.30	1.73	28.83
Diluent	24.98	0.09	0.12	0.31	0.16	3.48	29.14
Steam + Moisture	-	0.06	0.47	-	-	-	0.53
Total	48.32	1.64	2.10	0.77	0.46	5.21	58.50
Output							
Char + Diluent	39.97	0.49	0.65	0.64	0.34	5.68	47.77
Product Gas	1.12	0.34	0.08	0.00	0.01	-	1.55
Oils/Tars	5.69	0.53	0.68	0.06	0.05	0.05	7.06
Aqueous Condensate	0.00	0.28	2.18	0.00	0.00	-	2.46
Total	46.78	1.64	3.59	0.70	0.40	5.73	58.84
Output/Input	0.97	1.00	1.71	0.91	0.87	1.10	1.01
Diluent-Free Co-Product Yields, wt % Coal							
		dry			maf		
Char		65.3			61.3		
Gas		5.4			5.7		
Oils/Tars		24.5			25.9		
Water		6.7			7.1		
Total		101.9			100.0		

Table 3.56

Material Balances for the IGT Full-Stream Test IST-13
Coal: West Virginia, Fluidization Gas: 100% N_2, Temperature: 594°C

(Knight et al., December, 1990)

	Unadjusted						
	C	H	O	N	S	Ash	Total
Input							
Feed	24.56	1.48	1.75	0.49	0.34	1.61	30.23
Diluent	25.52	0.30	0.50	0.42	0.21	3.51	30.46
Steam + Moisture	-	0.08	0.62	-	-	-	0.70
Total	50.08	1.86	2.87	0.91	0.55	5.12	61.39
Output							
Char + Diluent	41.50	0.77	1.19	0.77	0.42	5.13	49.78
Product Gas	0.99	0.30	0.11	0.00	0.04	-	1.44
Oils/Tars	6.37	0.58	0.71	0.07	0.05	0.00	7.78
Aqueous Condensate	0.00	0.18	1.43	0.00	0.00	-	1.61
Total	48.86	1.83	3.44	0.84	0.51	5.13	60.61
Output/Input	0.98	0.98	1.20	0.92	0.93	1.00	0.99

	Adjusted						
	C	H	O	N	S	Ash	Total
Input							
Feed	24.56	1.48	1.75	0.49	0.34	1.61	30.23
Diluent	25.52	0.30	0.50	0.42	0.21	3.51	30.46
Steam + Moisture	-	0.08	0.62	-	-	-	0.70
Total	50.08	1.86	2.87	0.91	0.55	5.12	61.39
Output							
Char + Diluent	41.50	0.77	1.19	0.77	0.42	5.13	49.78
Product Gas	0.99	0.30	0.11	0.00	0.04	-	1.44
Oils/Tars	6.94	0.63	0.77	0.07	0.06	0.00	8.47
Aqueous Condensate	0.00	0.18	1.43	0.00	0.00	-	1.61
Total	49.43	1.88	3.50	0.84	0.52	5.13	61.30
Output/Input	0.99	1.01	1.22	0.92	0.95	1.00	1.00

Diluent-Free Co-Product Yields, wt % Coal		
	dry	maf
Char	64.2	62.2
Gas	4.8	5.0
Oils/Tars	28.0	29.6
Water	3.0	3.2
Total	100.0	100.0

Table 3.57

Material Balances for the IGT Full-Stream Test IST-14
Coal: West Virginia, Fluidization Gas: 100% N_2, Temperature: 691°C

(Knight et al., December, 1990)

	Unadjusted						
	C	H	O	N	S	Ash	Total
Input							
Feed	24.35	1.35	1.52	0.58	0.34	1.64	29.78
Diluent	25.24	0.30	0.40	0.41	0.21	3.59	30.15
Steam + Moisture	-	0.08	0.59	-	-	-	0.67
Total	49.59	1.73	2.51	0.99	0.55	5.23	60.60
Output							
Char + Diluent	43.46	0.56	0.73	0.91	0.40	5.45	51.51
Product Gas	1.68	0.56	0.37	0.00	0.04	-	2.65
Oils/Tars	3.77	0.30	0.28	0.04	0.03	0.00	4.42
Aqueous Condensate	0.00	0.12	0.91	0.00	0.00	-	1.03
Total	48.91	1.54	2.29	0.95	0.47	5.45	58.91
Output/Input	0.99	0.89	0.91	0.96	0.85	1.04	0.97

	Adjusted						
	C	H	O	N	S	Ash	Total
Input							
Feed	24.35	1.35	1.52	0.58	0.34	1.64	29.78
Diluent	25.24	0.30	0.40	0.41	0.21	3.59	30.15
Steam + Moisture	-	0.08	0.59	-	-	-	0.67
Total	49.59	1.73	2.51	0.99	0.55	5.23	60.60
Output							
Char + Diluent	43.46	0.56	0.73	0.91	0.40	5.45	51.51
Product Gas	1.68	0.56	0.37	0.00	0.04	-	2.65
Oils/Tars	4.71	0.38	0.36	0.06	0.03	0.00	5.54
Aqueous Condensate	0.00	0.12	0.91	0.00	0.00	-	1.03
Total	49.85	1.62	2.37	0.97	0.47	5.45	60.73
Output/Input	1.01	0.94	0.94	0.98	0.85	1.04	1.00

Diluent-Free Co-Product Yields, wt % Coal		
	dry	maf
Char	72.1	69.6
Gas	8.9	9.4
Oils/Tars	18.6	19.7
Water	1.2	1.3
Total	100.8	100.0

Table 3.58

Oils/Tars Elemental and Simulated Distillation Data For the IGT Task 4 PRU Tests, West Virginia Coal

(Knight et al., December, 1990)

Test Number	IST-10	IST-11	IST-12	IST-13	IST-14
Feed Coal	WV	WV	WV	WV	WV
Test Temperature, °C	598	599	594	584	691
Elemental Analysis of Oils and Tars[a]	dry wt %				
Ash	0.33	0.08	0.79	0.04	0.00
Carbon	79.93	80.18	79.36	81.50	84.14
Hydrogen	7.45	7.17	7.40	7.34	6.58
Nitrogen	0.73	0.90	0.89	0.89	1.20
Sulphur	0.88	0.53	0.83	0.72	0.67
Oxygen (by diff)	10.68	11.14	10.73	9.51	7.41
Total	100.00	100.00	100.00	100.00	100.00
H/C Atomic Ratio	1.11	1.07	1.11	1.07	0.93
Simulated Distillation by Gas Chromatography[b]					
Cumulative wt % Recovered	Boiling Point, °C				
5	181	ND	173	179	162
10	224	-	219	233	198
15	264	-	269	292	237
20	302	-	319	356	279
30	373	-	434	483	374
40	442	-	560	-	477
50	513	-	-	-	-
60	-	-	-	-	-
70	-	-	-	-	-
EP (end point)[c]	560	-	560	560	560
% Residue at EP	43.88	-	60.02	63.22	51.54

a) Determined by evaporation at 38°C, 15-20 mm Hg; light oils boiling below approximately 149°C are not included

b) Correction applied for heteroatom content of coal liquds, which is not accounted for in standard simulated distillation method for petroleum-based liquids

c) Characteristic of chromatographic column and method, not necessarily true end point of distillation

Table 3.59

Component Analysis of Full-Range Oils/Tars for the IGT PRU Task 4 Tests Using West Virginia Coal

(Knight et al., December, 1990)

Test Number	IST-10	IST-11	IST-12	IST-13	IST-14
Feed Coal	WV	WV	WV	WV	WV
Test Temperature, °C	598	599	594	584	691
Component	wt % of total oils/tars[a]				
Benzene	0.7	0.1	0.1	0.4	2.1
Toluene	0.5	0.5	0.4	0.4	1.3
Xylenes	0.4	0.5	0.3	0.5	0.8
Ethylbenzene	0.1	0.1	0.1	0.1	0.1
Indene	0.04	0.1	0.1	0.1	0.4
Styrene	0.1	0.1	0.1	0.1	0.2
Other Light Oils	9.1	ND	11.9	7.9	15.6
Total Light Oil[b]	10.9	ND	13.0	9.5	20.5
Phenol	0.4	0.1	0.3	0.3	0.5
Cresols	0.8	0.9	1.0	0.8	1.1
Xylenols	1.2	1.1	1.5	1.2	1.7
Naphthalene	0.1	0.2	0.4	0.1	0.3
Other Middle Oils	12.5	ND	8.9	8.2	9.8
Total Middle Oil[c]	15.0	ND	12.1	10.6	13.4
Heavy Oil[d]	11.9	ND	7.4	6.6	7.8
Pitch[e]	62.2	ND	67.5	73.3	58.3
Total Oils and Tars	100.0	ND	100.0	100.0	100.0

a) Includes light oils that are not included in the oils and tars of Table 3.58
b) Atmospheric boiling point < 182°C; estimated from simulated distillation data
c) Atmospheric boiling point 182 to 310°C
d) Atmospheric boiling point 310 to 399°C
e) Atmospheric boiling point > 399°C

Table 3.60

Gas Compositions for the IGT PRU Task 4 Test Using West Virginia Coal

(Knight et al., December, 1990)

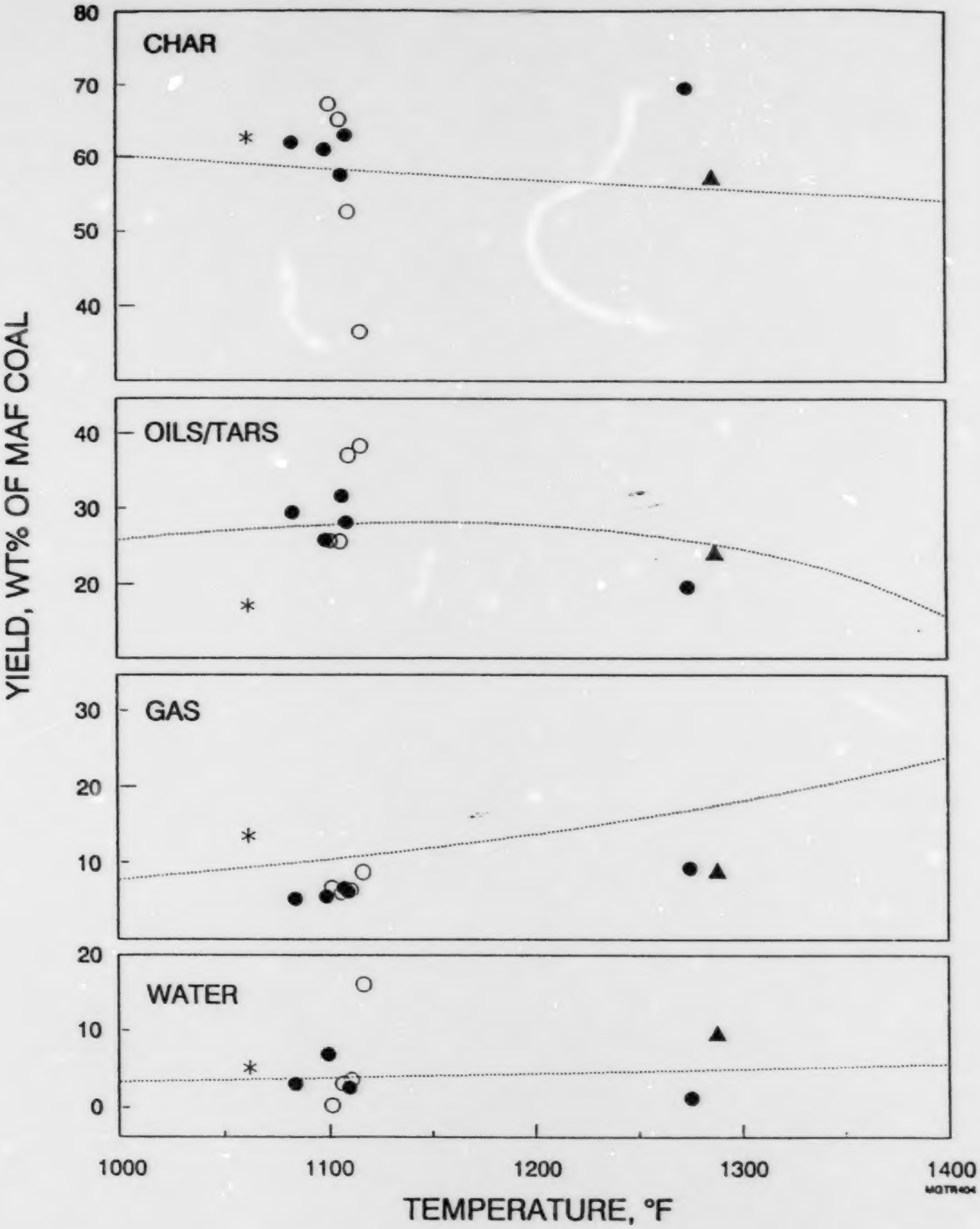
Test Number	IST-10	IST-11	IST-12	IST-13	IST-14
Test Temperature, °C	598	599	554	584	691
Component	Mol % in gas, nitrogen-free				
H_2	17.0	22.0	26.5	21.4	42.9
CO	5.6	5.3	1.9	3.1	7.1
CO_2	4.8	7.3	2.2	2.8	2.4
CH_4	53.0	47.2	48.5	52.4	34.5
C_2H_4	6.3	5.4	7.5	7.0	5.7
C_2H_6	7.4	7.4	7.2	6.6	3.4
C_3H_6	3.1	3.0	4.2	3.3	2.8
C_3H_8	1.4	1.5	1.6	2.0	0.6
H_2S	1.4	0.9	0.4	1.4	0.6
Total	100.0	100.0	100.0	100.0	100.0
Molecular Weight	18.9	18.7	16.7	17.6	13.8
High Heating Value, MJ/m^3	35.6	33.3	36.7	36.2	27.7

Table 3.61

Oils/Tars Analysis from the Full-Stream IGT PRU Task 4 Tests Using West Virginia Coal (Knight et al., December, 1990)

Test Number	IST-11	IST-12
Feed Coal	WV	WV
Test Temperature, °C	599	593
Total Sample	232.7	350.2
Solids	2.0	5.9
Distillate (overhead)	158.7	292.5
Tar (residue)	18.2	51.7
Loss	0.1	0.0
Adjusted Oils/Tars[a]	24.1	57.8
Elemental Analysis[b]	dry wt %	
Ash	0.0	ND
Carbon	87.0	ND
Hydrogen	7.4	ND
Nitrogen	0.1	ND
Sulfur	0.8	ND
Oxygen (by diff)	4.7	ND
Total	100.00	ND
H/C Atomic Ratio	1.01	ND
HHV, MJ/kg (calc)	38.4	ND
Component Analysis[c]	wt % of total oils/tars	
Benzene	0.01	0.10
Indene	0.06	0.08
Other < 182°C	0.73	0.50
Phenol[d]	0.08	0.19
o-Cresol	0.26	0.29
m,p-Cresols	0.38	0.41
C_2-phenols	1.01	1.30
Naphthalene	0.07	0.10
C_1-Naphthalenes	0.44	0.51
Other 182°-310°C	8.37	9.38
310°-399°C	24.93	20.02
> 399°C	63.66	67.12

a) Calculated from chromatographable tar in overhead + tar retained solvent
b) Distillation residue only
c) Toluene, xylenes, ethylbenzene and styrene are obscured solvent
d) Phenol in distillate is obscured by C_3-benzene peak in xylene value reported is from tar residue only



·········· Task 2 Data for Illinois No. 6 coal; Task 4 data: ○ Illinois No. 6 Coal; ● West Virginia Coal; ▲ Illinois Sesser Seam Coal; ✱ Rochelle Subbituminous Coal.

Figure 3.50

Co-product Yields From the IGT Task 2 and 4 PRU Tests

(Knight et al., December 1990)

The mild gasification pitch would require post-treatment to increase the aromaticity and remove heteroatoms in order to be acceptable as electrode binders. IGT suggests that on-stream upgrading consisting of exposure to condensables-laden product gas at 980°C would convert the oils/tars into forms favourable to binder pitch production. This treatment is expected to reduce the heteroatom content and increase the aromaticity, primarily by the de-alkalization of the aromatic centers in the high-boiling organic species.

A product development unit (PDU) has been designed to operate at 1 tonne/h and funding from the USDOE has been approved.

The PDU will use an integrated char recycle system that blends the recycled char with the coal feed and thereby remediates the caking problems that were seen in the PRU unit. A recycled coal derived liquid will be used as a quench liquid in the spray tower. A thermal cracker was added to allow processing of the pitch. Previous experience at IGT indicates that there should not be any problem with caking on the walls of an adiabatic unit.

The sulphur in form coke has not been addressed at this time. Research is continuing to examine methods of eliminating the problem. Hydrotreating is being considered.

The solids carry over into the quench tower has not been measured. This data will be collected during later tests. Some testing of hot gas filtration is being done. Barrier filters that would remove the particulates and leave the vapour phase are being considered. Co-generation of electricity is being considered for this mode of operation.

The utilization of the middle to light fractions as turbine fuel is being evaluated.

The use of the char as a boiler fuel is being looked at. The sulphur problem in the char needs to be addressed.

University of North Dakota

The objectives of the University of North Dakota program (Johnson et al., 1989) are to develop a process that will produce a low to medium heating value gas, a hydrocarbon condensate and a reactive char. The gas will be used to produce process heat, the condensate will be used directly or upgraded to diesel fuel, chemical feedstock, form coke binder or electrode binder pitch. The char will be used as a metallurgical coke or activated carbon. A 0.5 kg/h continuous fluid bed reactor has been operated.

A 45 kg/h fluidized carbonizer has been designed (Ness, 1989). The unit is designed to process non to moderately caking coals and is not suitable for processing N.S. Coals.

Western Research Institute

Merriam et al. (1989) have carried out either drying or partial gasification of Wyodak and Illinois No. 6 in a bench scale (4.5 kg/h) inclined fluidized bed reactor. Four products were targeted, (1) dirty pitch, (2) clean pitch, (3) heavy oil and (4) light oil.

The dirty pitch would be coated as dried coal prior to shipment; the clean pitch would be upgraded to electrode binder pitch; the heavy oil would be blended with diesel fuel and used in mining heavy equipment and the light oil would be sold as an octane enhancer. The philosophy of the process development was to produce value added products using technology that required the minimum of investment and operating costs.

The inclined fluidized bed reactor is operated at 3 and 0 degrees slope to give solids residence times of 5 and 10 minutes respectively. Fluidizing gas is introduced and product gases are drawn off along the length of the reactor so that there is no horizontal gas velocity. Process heat is applied externally (electrical) at up to 17.5 kJ/m^2 along the length of the reactor. The reactor is 1.5 m long, 15 cm wide and 19 cm high.

A water spray was used to remove coal dust from the exit gas during drying tests.

At the completion of a series of drying tests a second reactor was constructed to allow integrated drying/gasification operation. The limited capacity of the external electric heaters required that the fluidizing gas flow be kept small and hence the particle size be kept small. The gasifier is followed by an effluent gas cyclone and a two stage oil-condensing system. The heavy oil was condensed at 204°C and the remaining oil at 4°C. Tests were carried out with reactor temperatures from 635 to 660°C and average particle sizes from 5 to 10 mm.

Modifications were needed to the oil recovery system. These were to eliminate plugging by the condensation of the heavier products in sections of the system that were below their pour point. Some emission of vapour phase hydrocarbons could be seen with the modified condensation system.

Tests were carried out in the batch unit to examine the caking properties of Illinois No. 6. The heating rate and final temperature were varied to determine the conditions to use in the fluidized bed reactor.

Test results from the Illinois No. 6 will be discussed below.

A 5 cm ID fluidized bed was used to characterize the agglomeration characteristics of Illinois No. 6 coal. Coal was preheated in a CO_2 environment in a hopper. No agglomeration was seen after holding the coal for 15 min at 277°C. Severe agglomeration was seen after 10 min at 332°C.

The above data was used to set the temperature at which the coal would be dried prior to entering the gasifier. In all cases the partially caking Illinois No. 6 coal formed plugs in the reactor. Attempts to control temperature and heating rate only resulted in having the agglomerated mass form further down the length of the reactor.

It can be concluded that this reactor design is not suitable for the highly caking coals that occur in Nova Scotia.

4 PROCESS SELECTION AND PRELIMINARY EVALUATION

The most suitable process, in both the co-product and electrical generation formats, for further feasibility consideration will be selected. The selection process will put emphasis on; processes with proven performance and operability, suitability to handle Nova Scotia coals, flexibility to operate in a wide range of conditions and the associated environmental aspects of the process.

To facilitate the selection process, a quantitative comparative evaluation will be carried out separately for the co-product and electrical generation processes. Each of the relevant process steps will be given a weight from 1 to 10. Each process will have its status in the step in question scored from 1 to 10. The weighted score will be equal to the product of the score and the corresponding weight. The sum of the weighted scores for each process will be used to indicate the process that is considered to be most suitable for further evaluation.

4.1 Electrical Generation Process Selection

The initial evaluation will focus on technical aspects of the processes only. The operation of the electrical generation systems have been described in previous sections. Only the VEW and BC systems are considered to be developed to the point where further evaluation is justified. Based on those descriptions, the process steps shown in Table 4.1 are deemed the most significant and will be compared and given a quantitative evaluation.

Table 4.1

Weighted Comparison of the Process Steps in the Electrical Generation Processes

Process Step	Weight Step	---- VEW ----		--- BC ---	
		Score	Wt. Score	Score	Wt. Score
Gasifier	10	10	100	6	60
Char Combustor	7	6	42	8	56
Gas Turbine	7	7	42	8	56
Gas Particle Removal	6	7	42	7	42
Gas Phase Cleaning	5	7	35	6	30
Coal Handling	3	8	24	7	21
Total			291		265

Each category has been given a weight to reflect the significance that it has in relation to overall process. The following is the justification for applying a weight to each of the process steps.

The ability of a system to process Nova Scotia coals is primarily dependent upon the reactor design. The novel portion of the systems is the gasifier. The scale up potential of the reactor will be considered in this category. These considerations result in the reactor being given the maximum weight of 10.

The combustion of the char from the gasification step releases the remainder of the chemical energy in the coal to thermal energy. In the electrical generation process, the char combustor is also utilized to carry out some of the gas clean up (SO_2, NO_x, etc.). All of the systems considered in this final process selection utilize some form of fluidized bed combustor to burn the char. This technology has been developed extensively by others and could be utilized in conjunction with the gasifiers discussed here. The latter consideration results in the weight given to this step being a 7.

The release of the gas phase chemical energy to thermal energy and the reclaiming of that energy takes place in the gas combustor and gas turbine respectively. The gas turbine is the process step that is undergoing development. Corrosive gasses such as those containing chlorine and fluorine must be tolerated to some extent. This vital step in the energy conversion process has a weight of 7.

Particulates in the gas stream that are too large to follow the gas stream-line paths that pass through the turbine must be minimized in a removal step in order to avoid erosion. The operation of this step at high temperature and pressure allows the conversion efficiencies to remain high. This step receives a weight of 6.

Removal of gas contaminants is necessary in order to reduce corrosion in the gas turbine. This step receives a weight of 5.

The type of handling and conditioning of the coal prior to its entering the reactor is dependent upon the needs of the reactor. These types of systems are developed for numerous processes and should be easily available to the full scale versions of the partial gasification systems evaluated herein. This step receives a weight of 3.

4.1.1 Process Comparison

The review of the electrical generation process has revealed that only two systems are developed to a point where they need be considered in the final evaluation. The processes are those of British Coal and VEW.

The British Coal (BC) pressurized pilot scale gasifier is in the process of being commissioned. BC has extensive experience in the field of pressurized spouted bed combustion and atmospheric spouted bed gasification. These factors should enhance the process development. BC receives a score of 6.

The VEW gasifier has been in operation at 10 tonne/h for over six years. No significant operational problems have been encountered. The VEW gasifier receives a score of 10.

BC has carried out extensive testing of pressurized fluidized bed coal combustion and receives a score of 8.

The VEW process has relied upon the CFBC development of Steinmuller. The consideration of the utilization of other CFBC designs and the general state of development of the technology results in a score of 6.

BC's active role in the development of gas turbines for this application results in a score of 8.

VEW has not been as active as BC in the development of coal derived gas turbines. VEW does have 750 MWe of installed natural gas turbine fired co-generation capacity that does allow it to evaluate the co-generation mode of operation. VEW receives a score of 7.

High temperature and pressure particulate removal from coal derived gases is in the development stages. BC has an active program and receives a score of 7.

VEW has developed and operated a high temperature and pressure ESP based system that has operated on coal derived gas (score = 7).

4.1.2 Selection of a Process

The total weighted score for the BC system is 265 and for VEW is 292. This score reflects the slightly more advanced state of development of the VEW process. The larger scale, time of operation and associated technology related to coal gasification (natural gas/coal combined cycle) results in the selection of the VEW process for further evaluation.

4.1.3 Preliminary Evaluation of the Selected Process

The main process steps of the VEW process are shown in Figure 3.26. The process has been described in detail in Section 3.3.2.1. Figure 3.33 shows the configuration of the 750 MWe gas/coal combined cycle facility that is in operation at VEW's Dortmund, Germany facility. Some similarities to a coal partial gasification combined cycle facility are seen. Coal derived gas from the coal gasifier would be cleaned, fired and exhausted into a gas turbine. The gas turbine would be exhausted into a CFBC where it would join the hot char from the gasifier.

- Material Balances

Figure 4.1 shows some of the mass flows for a 108 tonne/h of coal facility that is shown to be generating 379 MWe.

Based on the data provided the following calculations have been made. The analyses of Prince coal in Tables 3.3 to 3.6 have been used. The Ca/S molar ratio has been assumed to be 2.5 for the overall process. The sulphur capture rate is assumed to be 95%. The excess air is assumed to be 20%. All of the carbon in the sorbent is assumed to be released as CO_2.

Based on the above and the data in Figure 4.1 the emission rates have been determined to be as follows; CO_2, 2,130 g/MJe and SO_2, 0.294 g/MJe.

The effect of the coal conversion efficiency and the volatile content of the coal on the lower heating value of the gas produced in the gasifier is shown in Figure 4.2. Analysis of coals used in the VEW 10 t/h gasifier are shown in Table 4.2.

Table 4.2

Analysis of Coals Used in the VEW 10 t/h Partial Gasification Process

	AV	AC	WF	IBB	MO-E	MO-G
Volatile Matter Wt. % (maf)	30.6	26.8	22.1	6.4	15.1	29.8
Ash Wt. % (maf)	6.3	7.2	7.4	9.0	4.6	10.0
C Wt. % (maf)	80.2	81.0	81.8	83.9	85.6	73.5
H Wt. % (maf)	5.1	5.0	4.6	3.2	4.2	4.8
O Wt. % (maf)	5.8	4.0	3.6	1.8	3.1	8.4
N Wt. % (maf)	1.6	1.6	1.4	1.1	1.6	1.6
S Wt. % (maf)	1.04	1.01	1.09	0.8	0.8	1.5
Cl Wt. % (maf)	0.15	0.16	0.15	0.19	0.10	0.21
F Wt. % (maf)	0.007	0.008	0.009	0.011	0.008	0.008
Heating Value MJ/kg (mf)	32.1	32.2	33.0	32.0	34.1	30.1

Coals

AV - Auguste Viktoria
AC - Achenbach
WF - Westfalen
IBB - Ibbenbüren
MO-E - Monopol Esskohle
MO-G - Monopol Gaskohle

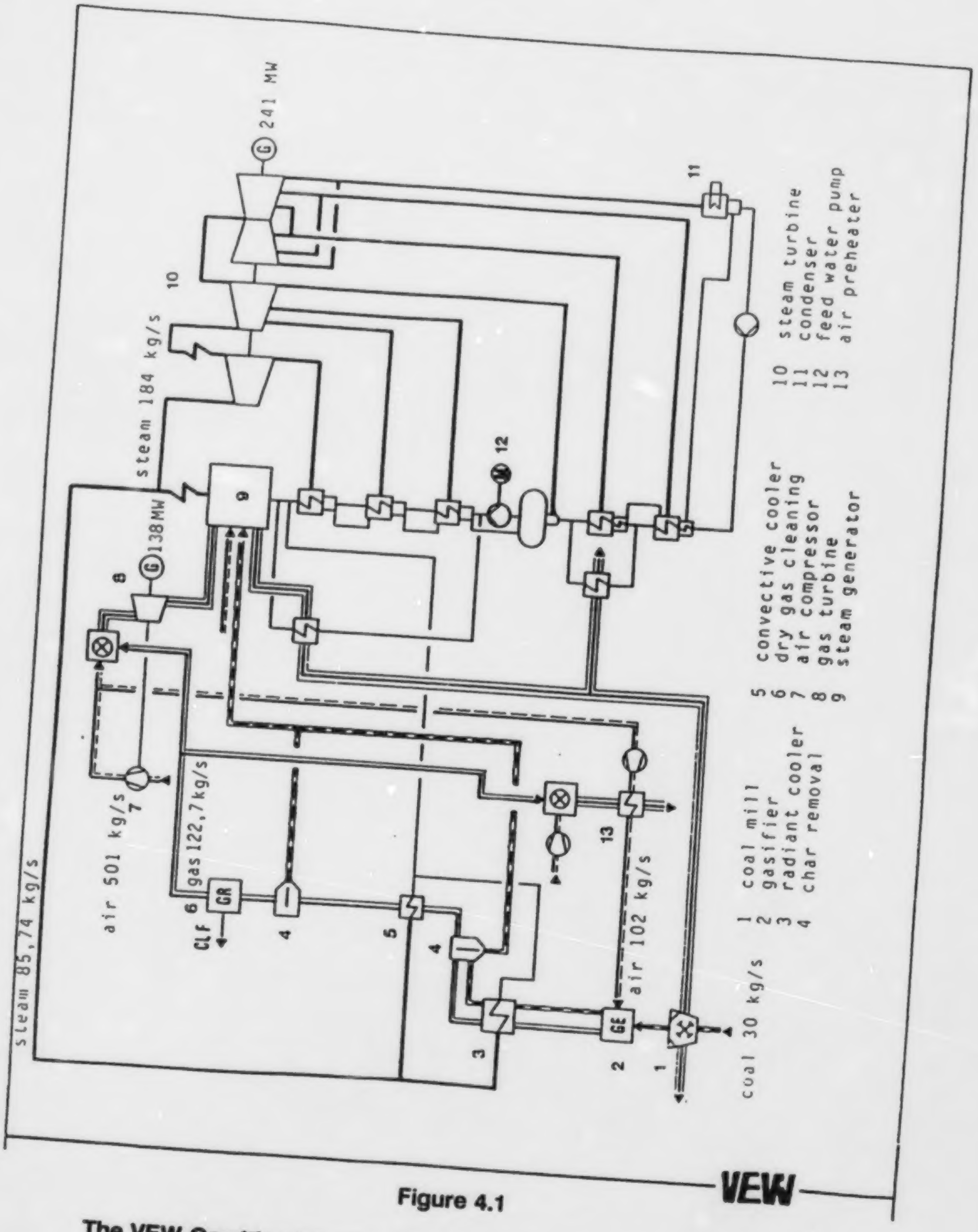

steam 85,74 kg/s
air 501 kg/s
gas 122,7 kg/s
steam 184 kg/s
241 MW
138 MW
coal 30 kg/s
air 102 kg/s
CLF
GR
GE
1
2
3
4
5
6
7
8
9
10
11
12
13
1 coal mill
2 gasifier
3 radiant cooler
4 char removal
5 convective cooler
6 dry gas cleaning
7 air compressor
8 gas turbine
9 steam generator
10 steam turbine
11 condenser
12 feed water pump
13 air preheater
VEW

Figure 4.1

The VEW Combined Cycle with Integrated Partial Coal Gasification
Process Flow Diagram

Typical gasifier gas composition is shown in Table 4.3.

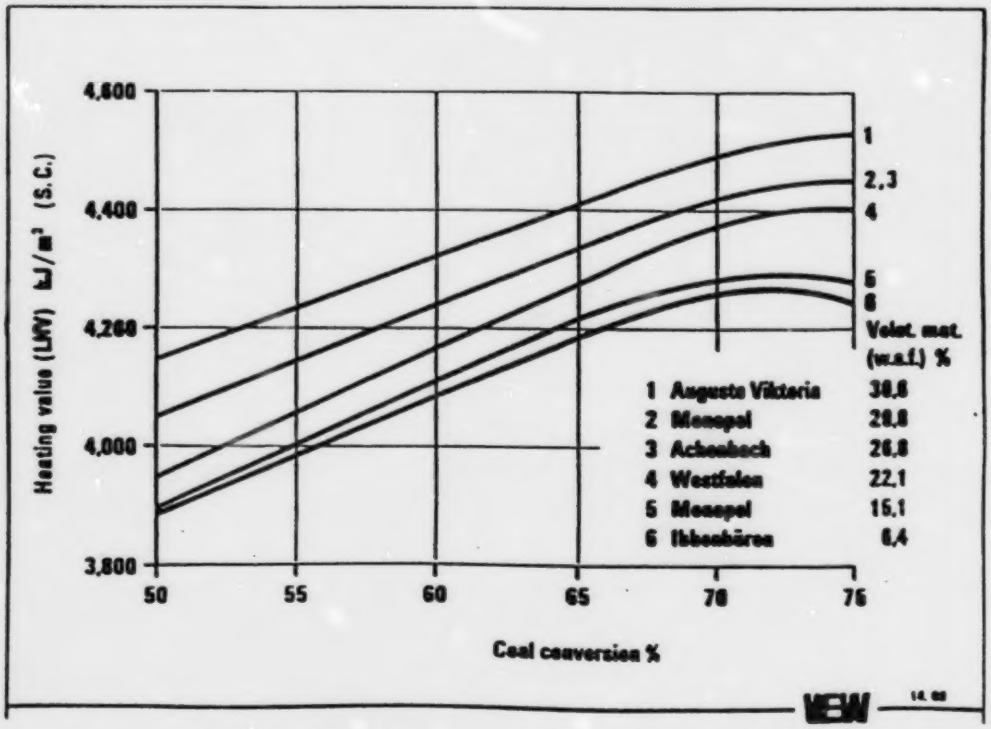
Table 4.3

Typical Composition of Gas Entering the Gas Turbine Combustor In the VEW 10 t/h Partial Gasification Process

Gas	Composition Volume %
CO	22
H_2	14
CO_2	4
H_2O	5
N_2	55

A lower heating value of approximately 4.4 MJ/Nm^3 should be achievable with a Nova Scotia coal. The projected LHV for the gas is arrived at after comparative analysis of the Nova Scotia coal analysis (Tables 3.3 to 3.10) and the German coals used in the VEW reactor (Table 4.2) and the results shown in Figure 4.2. This lower heating value would allow a high portion of the heating value of the parent coal to be utilized in the gas turbine. After combination with additional air the O_2 concentrations in the turbine exhaust gas are expected to be in the 14% range.

Figure 4.3 shows the effect of gas turbine inlet temperature and gas cleaning method on the process efficiency at 70% coal conversion. Gas turbine materials may limit the inlet temperature. Dry gas cleaning improves the efficiency by approximately 1%. The higher steam pressure curve (230 bar) would require supplementary firing in the steam generator. Heating of the compressed air (destined for the gasifier) from the gas turbine in the steam generator will further reduce the energy input needed from the coal. This would allow a further increase in the efficiency of approximately 2%.



10 t/h Plant / Gas-Heating value

Figure 4.2

The Effect of Coal Conversion and Coal Volatile Content on the LHV of the Gas from the VEW Gasifier

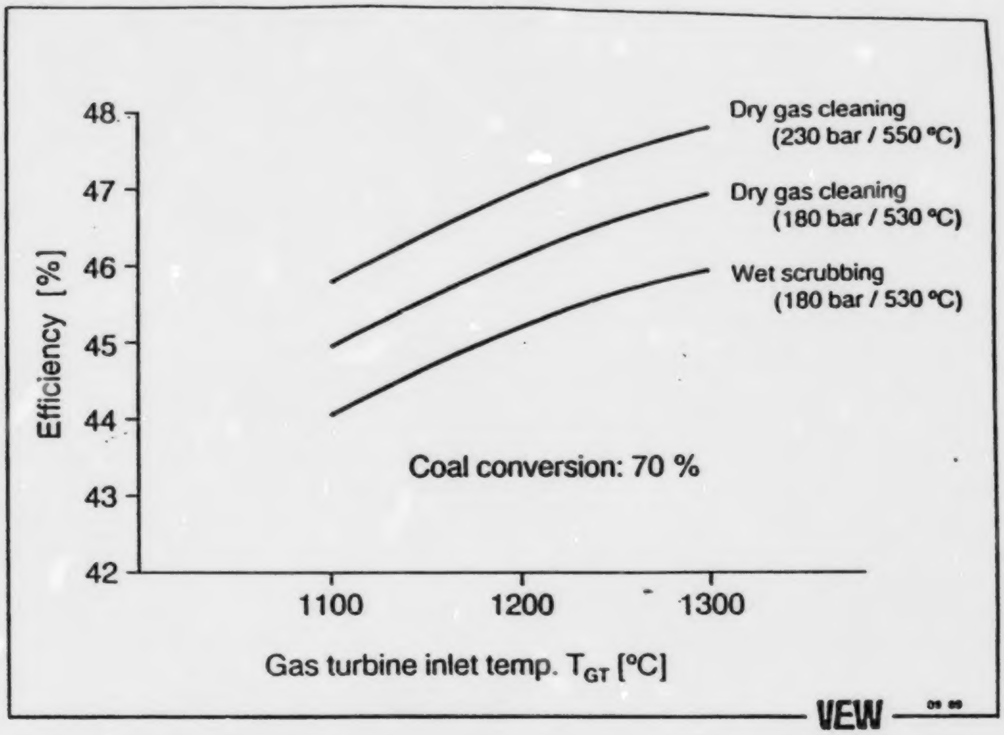


Figure 4.3

Efficiencies for Different Process Modifications, VEW

VEW has made recommendations for systems configured to better follow load variations. Figure 4.4 shows the efficiency of the process as the percent of full load is varied. A nearly constant efficiency is achieved by varying the coal conversion and using a system with two gas turbines. At reduced loads, the coal conversion is increased which keeps the LHV of gas constant. This allows the operation of the gas turbine at a constant inlet temperature and output energy level. The overall reduction in electrical output is as a result of the reduced output of the steam turbine alone. A turndown from 100% of full load to 58% of full load is achieved (at a relatively constant efficiency) as a result of increasing the coal conversion in the gasifier from 60 to 75%. A further turndown requires the shutting down of one of the gas turbines and reduction of the conversion to 60% again. This mode of operation allows operation of the gas turbines at nearly constant compressor throughput and therefore nearly constant efficiency.

Figure 4.5 has been prepared by Weinzierl (1985) to present the effect of process design and operational conditions on cycle efficiency. The cycle efficiency, η, is plotted versus the ratio of gas turbine to steam turbine output (NGT/NDT). Examples of total gasification (O_2 blown) and partial gasification are shown, each with gas turbine inlet temperatures of 950 and 1250°C. Both single and dual pressure range steam cycles are shown for the total gasification processes. As the NGT/NDT ratio increases, the gas turbine exhaust temperature increases. A second steam pressure condition allows the capture of the waste heat in the higher temperature turbine exhaust gas. The partial gasification process achieves the same efficiency as the total gasification at a lower NGT/NDT value due to the operation of the partial gasification process as a supplemental fired system. Operating at lower NGT/NDT ratios reduces the dependence of the cycle efficiency on the gas turbine inlet temperature. The partial gasification curves are prepared assuming steam properties of 18.2 MPa and 530°C with a condenser pressure of 70 kPa. The partial gasification curve labelled "Coal gas design" uses wet gas cleaning and the curve labelled "Flue gas design" utilizes dry/hot gas cleanup. A curve which shows a further increase in efficiency is presented for steam at 550°C and at supercritical pressure. Supercritical steam properties can also be achieved with a total gasification process but at the expense of the gas turbine capacity. The gain in efficiency achieved in the steam cycle is lost in the overall efficiency due to the decrease in the NGT/NDT ratio as a result of the combustion of a lower fraction of the coal derived gas in the gas turbine.

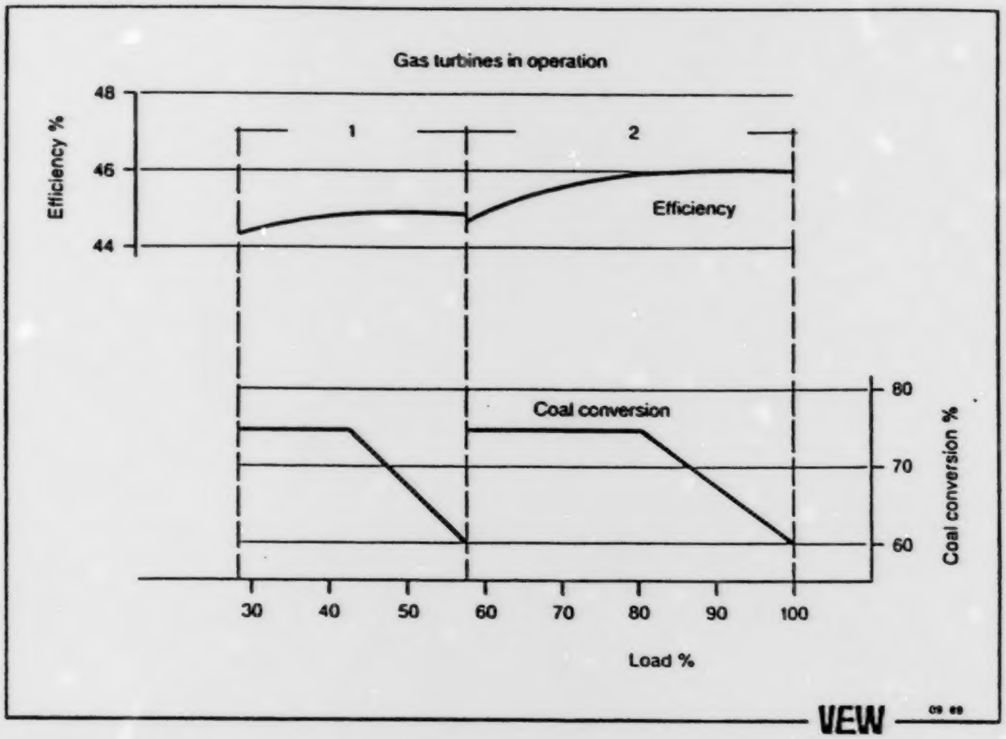


Figure 4.4

Efficiency of the VEW Partial Gasification Process As a Function of the Percent Full Load

(Weinzierl, 1989)

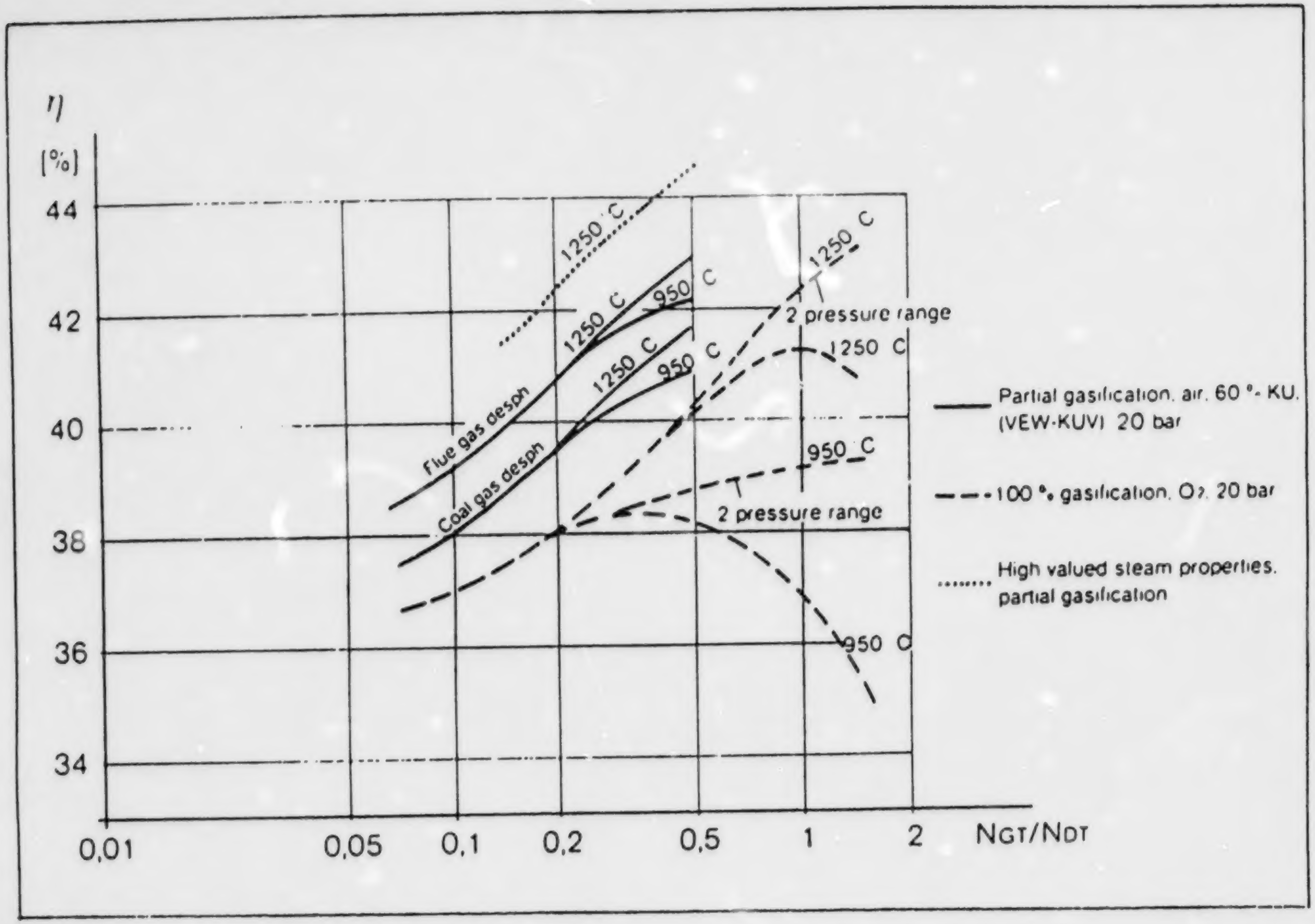


Figure 4.5

VEW Gas/Steam Turbine Optimization Curves

(Weinzierl, 1985)

4.2 Co-Product Process Selection

The initial evaluation will focus on technical aspects of the processes only. The operation of the co-product systems have been described in previous sections. Based on those descriptions, the following process steps are deemed the most significant and will be compared and given a quantitative evaluation:

Category	Weight
Reactor	10
Product Separation	8
Liquid Product Processing	7
Solid Product Processing	6
Emission Control	5
Coal Feed Handling and Preparation	4

Each category has been given a weight to reflect the significance that it has in relation to overall process.

The ability of a system to process Nova Scotia coals is primarily dependent upon the reactor design. The processability and subsequent market value of the products is dependant upon (with coal feed stock being a constant) the reactors design and operational conditions. The scale up potential of the reactor will be considered in this category. These factors result in the reactor being given the maximum weight of 10.

Once the reactor has preformed its function, the system must be capable of retrieving the products. Product separation involves the separation of the three phases that the products may be in, gas, liquid and solid. This is a critical portion of the process and receives a weight of 8.

The processing of the products so that they are in a form that is either marketable or can be handled prior to further upgrading is in some cases necessary. Typically, the liquid products from gasification of coal are in a more unstable form than the gas or solid phases. Heteroatoms can cause re-polymerization and gumming or sedimentation. Liquid product processing receives a weight of 7.

The solid products are relatively stable and the processing would be to add value to form products such as form coke or activated carbon. The step receives a weight of 6.

All of the process steps must be designed and operated so as to minimize the emissions to the environment. This consideration receives a weight of 5.

The handling and conditioning of the coal prior to its entering the reactor is an important step. These types of systems are developed for numerous processes and should be easily available to the full scale versions of the partial gasification systems evaluated herein. This step receives a weight of 4.

4.2.1 Process Comparison

Three currently active processes have been reviewed in this document that would be technically suitable to carry out partial gasification of Nova Scotia coals. They are the British Coal (BC) spouted fluidized bed system, the Coal Technology Corporation (CTC) screw reactor system and the Institute of Gas Technology (IGT) bubbling/entrained fluidized bed system.

- Coal Feed Handling and Preparation

Table 4.4 summarizes the coal handling systems discussed below.

Table 4.4

Summary of the Co-Product Partial Gasification Processes Coal Handling Systems

Process	Scale of Operation (kg/h)	Coal Size Handled (mm)
British Coal	500	0.6 & 15-25
Coal Technology Corporation	454	–
Institute of Gas Technology	45	0.4 - 1.7

The earlier work on the BC process was carried out on crushed coal with a top size of 0.6 mm. Later work was done on a specific coal feed that is available in the United Kingdom, "singles" in the size range of 12 - 25 mm. The feed system is relatively simple (Figure 3.20) utilizing hoppers and mechanical conveying. The system has been operated for extended periods at feed rates up to 500 kg/h. This stage of development results in a score of 9.

The CTC is in the process of scaling up its feed system in a 454 kg/h scale process. Details on the feed system are not available. It should utilize conventional technology and scale up should not pose any problems although demonstration of operational success is needed. The lack of data and the stage of development result in a score of 7.

The IGT process (Figure 3.44) has been operated at 45 kg/h to feed coal to the 69 kPa pressure reactor. The coal and coke breeze sizes that were fed ranged from 0.4 to 1.7 mm (Table 3.45). Pressure locks and pneumatic transport are used to move the coal into the reactor. Some problems have been encountered in the feeding of caking coals that have been addressed. This status results in a score of 7.

Table 4.5 summarizes the Co-product system reactor designs.

Table 4.5

Summary of the Co-Product Partial Gasification Processes Reactor Systems

Process	Scale (kg/h)	Reactor Design	Pressure (kPa)
British Coal	500	Spouted/ Fluidized bed	0
Coal Technology Corporation	454	Screw	0
Institute of Gas Technology	45	Bubbling/ Circulating Fluidized Bed	69

The BC Spouted/Fluidized bed reactor (Figure 3.19) has operated in the 500 kg/h range for more than 700 hours in the low coal conversion variant (65%) and achieved successful performance. The high coal conversion variant (90%) has operated for more than 1750 hours. No significant operational problems have been reported. A utilization for the ungasified carbon in the low conversion variant has not been established. Scale up of this design should not pose any significant problems. The scale, duration and successful operation achieved with the BC reactor results in it achieving the maximum score of 10.

The CTC screw reactor (Figure 3.39) is in the process of scaling up to 454 kg/h. Operational data with this scale of reactor is not available at the time of preparation of this report. This reactor system must be considered to be unproven at this scale. The mode of handling of swelling and caking coal as we have in Nova Scotia in a reactor of this design could prove very troublesome, especially at larger scales. Due to the lack of available data and scale of operation of the CTC system the score is 4.

The IGT bubbling/circulating fluidized bed reactor (Figure 3.44) has been operated at 45 kg/h. Knight et al. (July and December, 1990) has reported that the problems associated with processing caking coals have been addressed to some degree. It is expected that larger scale reactors will have higher char inventories which will negate the need to recycle char. This reactor has the most extensive reporting of operational conditions of the three reactors discussed in this document. The reactor has not been operated with integrated process generated char recycle. Char from previous tests has been used to simulate this operation. Some process development will be needed in this area. Considerations of the scale of operation and the development work needed results in the IGT reactor scoring 6.

- Product Separation

The BC system (Figure 3.20) uses primary and secondary cyclones, and in some cases, a bag filter to remove particulates from the reactor off gas. Levels of particulates in the cleaned gas ranged from 300 to 960 mg/m^3 using cyclones only. The addition of a bag filter reduces the loading to < 10 mg/m^3. All gas phase material would be sent to a combustor. The particulate levels achieved using a bag filter would be acceptable for use in most gas combustion systems. Hot gas filters are under development. The simplicity of the system and the duration of operation of the BC process results in a score of 8.

The CTC (Figure 3.39) utilizes cyclones, scrubbers and condensers to separate the particulate material that has been carried over with the sweep gas. Very little data is available on the actual design or operation of the 45 or 454 kg/h scale product separation systems. CTC reports that gas produced in a commercial operation would be used to supply process heat. Present operation sees the non-condensable gases flared. Solids concentrations in the condensable liquids recovered from the gasification operation have not been reported. The lack of data and the present stage of development results in a score of 5.

The IGT product separation PRU full-stream condensate system is shown in Figure 3.48. Two stages of cyclones are electrically heated to above the tar condensation temperature. It will be presumed that the particulate material not collected in the cyclones is removed in the bottom drain of the quench tower. A larger scale system would utilize some other form of maintaining the cyclone above the tar condensation temperature. The IGT 45 kg/h PRU was operated using a spray tower to collect the middle and heavy distillate fraction of the condensable hydrocarbons. In the PRU system, the light fraction was vented to a combustion system. Development work will be needed to collect this fraction. Modifications (as discussed earlier) were made to the PRU full stream spray tower to improve it's efficiency. The commercial IGT plant would utilize the gas to produce process heat. The development needed results in a score of 6.

- Liquid Product Processing

The BC system produces a low tar gas and therefore does not have liquid product processing. As above the BC system will receive the mean of the score that the other two systems receive in this category (score = 7).

The CTC system has reported extensive upgrading of liquid products. The upgrading has been carried out on coal derived liquids that were not generated in their process. This allows CTC to gain experience in the area, but the particulars that may be associated with the upgrading of the liquids generated from the CTC process are not addressed. The CTC liquid product process receives a score of 8.

The IGT liquid products are intended to be pitch or chemical feedstocks. The pitch produced from the IGT process would require post-treatment to increase the aromaticity and remove heteroatoms. The IGT process needs further development in the area of liquid product processing and receives a score of 6 in this area.

- Solid Product Processing

The BC system has been targeted for use in brick making. The solid residue would be added to the brick material. This very limited application would leave up to 40% of the coal unused in the low conversion variant and 15% in the high conversion variant. As a result of the above the BC system receives a score of 0 in this category.

CTC has used a modified coke testing procedure to characterize coke made in a small scale apparatus from products from their process. The products produced were acceptable according to the modified test procedure. The production of form coke is carried out on commercial scale in other applications which should allow the utilization of products from the CTC process in a commercial operation. The CTC process receives a score of 6 in this category.

The IGT process reports the successful production of form coke, smokeless fuel and activated carbon. As with the CTC system, the utilization of products from the IGT system in commercial scale operations should be feasible. A score of 6 is given to CTC.

- Emission Control

The BC system would have to deal with particulate in the gas stream. Operation using a bag filter has been reported but this would add a thermodynamic penalty due to the gas cooling needed. Particulate emissions that reach a gas combustor may be burnt in it. All of the gas phase material would be burned in normal applications. The utilization of the 15 to 40% of the unconverted carbon needs to be addressed. The BC system receives a score of 2 due to the last consideration.

The CTC system is proposed to operate such that all of the products would be utilized. The collection of fugitive particulate material is not discussed in the CTC information that is available. The lack of information in this area results in a score of 6.

The IGT system is as the CTC system discussed and receives a score of 6 also.

Table 4.6 summarizes the weights and scores that the three co-product processes received in each of the categories of evaluation. This evaluation is based on technical aspects only. The variation in the total weighted scores is low enough that further evaluation is necessary. The marketability of the products must be explored to allow the selection of a process to be made for subsequent evaluation.

Table 4.6

Process Selection Scores for the Co-Product Processes

Process Step	Weight	--- BC ---		--- CTC ---		--- IGT ---	
		Score	Wt.Score	Score	Wt.Score	Score	Wt.Score
Coal Handling	4	9	36	7	28	7	28
Reactor	10	10	100	4	40	6	60
Product Separation	8	8	64	5	40	6	48
Solid Product Process	6	0	0	6	36	6	36
Liquid Product Process	7	7	49	8	56	6	42
Emission Control	5	2	10	6	30	6	30
Total			259		230		244

- Marketability of Products

This section of the report must be somewhat speculative due to the complexity of the issue. Numerous factors are at play for each of the products.

Although the BC system has been developed and operated at the largest scale, the lack of a use for the unconverted carbon is a significant drawback, especially for the low conversion variant. The only product that would be available for marketing is the low to medium heating value gas. Only the BC system targets the marketing of gas. In Nova Scotia the utilization of gas on an industrial scale is limited. The introduction of pipeline gas to the region would result in a market being established but would also result in stiff competition.

The CTC system will produce form coke and transportation fuels with the coal derived gas being used on site. Recent activity (Price 1991) in Nova Scotia indicates that an export market exists for metallurgical coke. The acceptance of the coal derived fuels into the market place will be dependant upon a number of factors: price (which will be to some extent a function of the public utilities board and taxation), consumer attitude, environmental legislation regarding the exposure to the aromatic components in the fuel and cooperation with local producers and suppliers (i.e. access to distribution and retailing facilities). The price of conventional sources cannot be predicted. The lack of data makes the evaluation of the CTC process somewhat speculative.

The IGT system will produce a gas product that will be mostly utilized on site in a commercial operation, three options for the solid products: form coke, smokeless fuel and activated carbon and liquid chemicals to be used to make electrode binder pitch or chemical feedstock. IGT has identified markets in the United States in the millions of tonnes per year range for the solid products (30 x 10^6 tonne/yr for coke, 50 to 200 x 10^6 tonne/yr for smokeless fuel and 1 to 2 x 10^6 tonne/yr for activated char). The versatility of producing a variety of solid products would enhance the viability of the process.

4.2.2 Selection of a Process

Consideration of the marketability of the products and the process evaluation results in the selection of the IGT process for further evaluation.

4.2.3 Preliminary Evaluation of the Selected Process

The yields of the IGT process have been presented in Tables 3.50, 3.53, 3.54, 3.55, 3.56 and 3.57. Table 3.50 is based on the slipstream samples taken from the gas stream. The procedures that have been used to arrive at the mass balance data have been described in Section 1.3.2.2. Both of the tests, MG-28 and MG-30, used coke as a dilatant to prevent agglomeration of the caking coal. Table 4.7 summarizes the operational conditions and the yield data while processing West Virginia coal for the IGT process. Figure 4.6 shows the yield data for the IGT process. The data is presented in order of increasing reactor temperature.

Table 4.7

Summary of Operational Conditions and Yield Data for the IGT PRU System Operating on West Virginia Coal

Test ID	Temp. (°C)	Total Feed Rate (kg/h)	Coal Feed Rate (kg/h)	FB Height (cm)	Solids Res Time (min)	Yield Data (maf)			Oils (%)	Pitch (%)	Oils/Pitch Ratio	Water (%)
						Char (%)	Gas (%)	Oils & Tars (%)				
MG-28	621	23.5	7.8	43	15	69.4	8.8	22.8	7.6	15.2	0.502	-1
MG-30	566	30.5	10.2	114	17	73.4	5.7	21.4	7.3	14.1	0.520	-0.5
IST-10	597	24.3	12.2	58	13	57.8	6.9	31.8	12.0	19.8	0.608	3.5
IST-11	599	33.6	16.8	69	9	63.2	5.8	28.3				2.7
IST-12	593	28.6	14.3	49	10	61.3	5.7	25.9	8.4	17.5	0.481	7.1
IST-13	584	27.8	13.9	70	11	62.2	5.0	29.6	7.9	21.7	0.364	3.2
IST-14	691	27.5	13.8	49	9	69.6	9.4	19.7	8.2	11.5	0.715	1.3

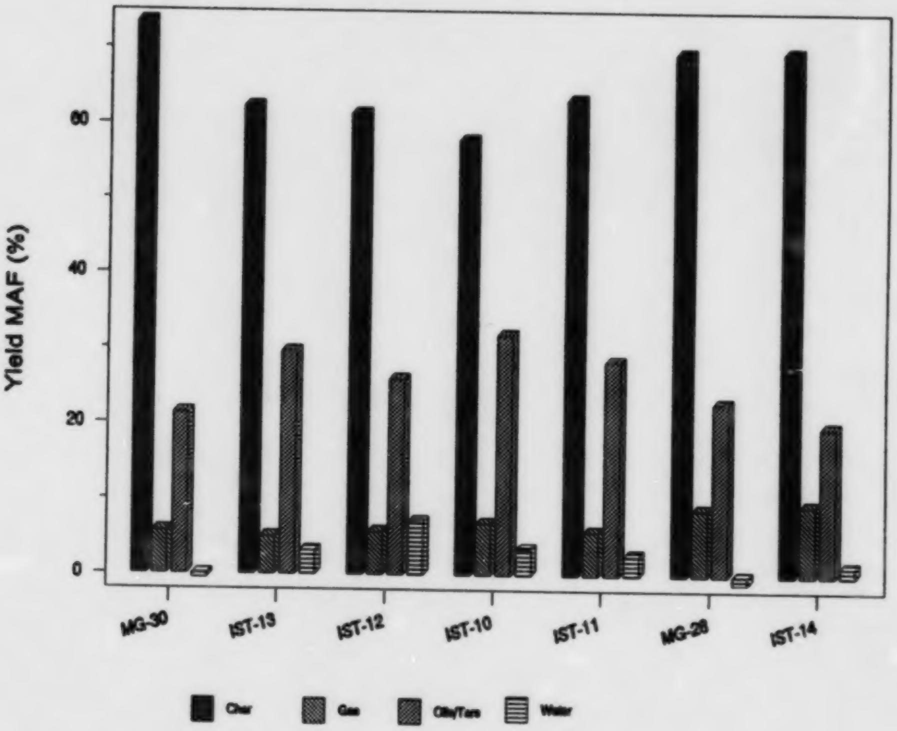


Figure 4.6

Yields from the IGT PRU System While Processing West Virginia Coal

Char yields varied from 58 to 74%. The operational condition that was varied the most between these two tests was the fluid bed height (58 to 114 cm). Gas yields varied from 5 to 9.4% while the reactor temperature was varied from 584 to 691°C respectively. The oils/tars yield varied from 19.7 to 31.8%. The lower temperature and longer bed residence times would have contributed most to the higher oils/tars yields. Variations in the yields should be viewed in light of the different adjustment procedures used in the MG and IST runs as discussed in the text above.

A commercial operation would see the utilization of recycled coal derived char as both a diluent and a heat source. The chemical and pore structure of the char may have some effect on the relative amounts of the gas, liquid and solid phase yields. Comparative evaluation of the yield data from the runs using coke or char as a dilatant is difficult due to the variation of other operational parameters such as the solids residence time. In addition, the composition of the gas and liquid phases may vary due the change in the catalytic effect of char as apposed to that of the coke.

N_2 was used as a fluidizing gas in the PRU tests. A commercial operation would use recycled coal derived gas. The change in the chemical composition of the fluidizing gas will have some effect on the composition of the gas and liquid phases produced.

The yield and mass flow data from the processing of a Nova Scotia coal would be within the range of data presented for the West Virginia coal.

5 VEW PARTIAL GASIFICATION PROCESS; PROCESS DESCRIPTION AND ECONOMICS

An economic evaluation of a conceptual application of the VEW partial gasification process to the generation of electricity utilizing a Nova Scotia coal has been carried out. The conceptual generation scheme has been sited at the Nova Scotia Power Corporation's (NSPC) Point Aconi Circulating Fluidized Bed Combustion (CFBC) location. The VEW gasifier is combined with a CFBC to generate 284 MWe (net). NSPC cost data has been utilized in evaluating the CFBC related capital, operating and maintenance costs. A study carried out by British Coal (Blamire, 1991) to evaluate the British Coal Topping Cycle sited at Point Aconi was utilized in estimating additional related costs.

5.1 Process Description

This portion of the report describes the various sections of the VEW based partial gasification system to generate electricity. The electrical generation plant descriptions are divided into gasification, combustion and electricity generation sections.

5.1.1 Gasification Sections

The VEW gasification system has been described in detail in previous sections in this report. A single 2.65 m internal diameter gasifier operating at 2 MPa and 1600°C will accept 75 t/h of Prince coal (see Tables 3.3 to 3.6) to produce 284 MWe (net). Weinzierl 1985 has reported that a commercial gasifier would have a capacity of approximately 70 t/h. The gasifier will convert 60% (wt basis) to gas with the remaining going to the CFBC as hot char. The overall cycle efficiency (chemical energy, HHV, in coal to electrical energy to grid) is assumed to be 44%. A HHV of 31 MJ/kg has been used for the coal (see Table 3.3).

The coal sizing and handling system would be capable of handling coal (output) at 21 kg/s.

The air supply to the gasifier would have a capacity of approximately 60 kg/s. It is assumed that a stand alone air preheater will not be necessary as the air will be heated during the compression step.

Chlorine and fluorine will be removed from the hot gas prior to entering the gas turbine combustor. It is envisaged that a dry gas cleaning system will be used to optimize the system efficiency.

Table 5.1 shows the operational conditions and mass flows for the VEW gasification system.

Table 5.1

Mass Flow Rates and Operating Systems for the Conceptual VEW System Located at Point Aconi

Gasifier Section	Specifications
Coal Milling	21 kg/s to < 90 μm
Coal Pressurization	21 kg/s to 2 MPa
Air Compressor	350 kg/s to 2 MPa
Gasifier and Radiant Cooler	86 kg/s steam at 18.2 MPa & 535°C
Hot Char Transfer to CFBC	8.3 kg/s at 800°C
Hot Gas Particulate Removal	12.5 kg/s gas, 10 mg/Nm^3 exit loading
Chlorine & Fluorine Removal (hot gas)	75 g/s Cl & 4.5 g/s F at 90% efficiency
Gas Turbine/Electrical Generator	12.5 kg/s of gas with LHV 4.4 MJ/m^3, 103 MWe

5.1.2 Combustion Sections

Approximately 8.4 kg/s of char at 800°C will be fed to the CFBC similar in design to that under construction at Point Aconi for NSPC. The majority of the char will be collected in the cyclone separator after the convective cooler. Steam generated in the CFBC from the combustion of the hot char will contribute approximately 83 MWe to the plant output.

5.1.3 Electricity Generation Sections

The radiant cooling section of the gasifier will produce approximately 86 kg/s steam at 18.2 MPa and 535°C. This steam will be combined with that produced from the combustion of the hot char in the CFBC and enter the steam turbine which is rated at 181 MWe. Additional heat recovery is carried out as the char is cooled further in the convective section to 800°C.

It is proposed that this conceptual system will be operated as a base load plant. This mode of operation will negate the necessity of operating two gas turbines. A single gas turbine rated at 103 MWe will be used. This will result in a ratio of gas turbine to steam turbine output of 0.57. Referring to Figure 4.5, this will ensure operation in a region of high efficiency.

5.2 Economic Analysis

It is the objective of this analysis to evaluate the economics of generating electricity by using the VEW partial gasification system in conjunction with a CFBC.

5.2.1 Capital Costs

Detailed cost information has been generated by NSPC for the Point Aconi 165 MWe CFBC generating station. Some of this data has been utilized in this study to arrive at costs for the CFBC, steam generator and other associated equipment. The NSPC data indicates that a significant reduction in the capital costs are seen due to the utilization of common facilities. This advantage has been used in this analysis. In addition, information generated for NSPC by British Coal (Blamire, 1991) has been used to estimate capital costs.

Power Plant and Gasifier Investment

The following procedure has been used to arrive at an estimate for the capital costs of the power plant.

The British Coal study has presented costs for a 350 MWe topping cycle system sited at Point Aconi. Data made available from NSPC for the Point Aconi CFBC has been used. The capital cost for the facilities that would be similar to the Aconi 165 MWe CFBC and used in the 284 MWe VEW gasifier/CFBC facility have been scaled using the following equation:

$$\frac{\$\ unit\ 1}{\$\ unit\ 2} = \left(\frac{MW1}{MW2}\right)^{0.7}$$

where \$1 and \$2 are the costs associated with units one and two respectively and MW1 and MW2 are the generation capacities of units 1 and 2 respectively.

All costs have been inflated at 5% to mid 1991 Canadian dollars.

Table 5.2 outlines the plant cost information used to arrive at a total plant investment.

Table 5.2

Plant Cost Summary for the Conceptual VEW System Located at Point Aconi (all values are in mid 1991 CDN$ *1E6)

Process Facilities	
Coal Reception and Storage	$0.0
Sorbent Reception and Storage	$0.0
Sand Handling	$0.8
Cumbustor Feed	$1.8
Feed Handling	$6.6
Gasification	$13.1
Hot Gas Cleaning (Gasifier)	$13.2
Gasification Air Compressor	$4.9
Gas Turbine	$38.3
CFB Combustor/Cyclones	$36.2
Clean Gas Boiler	$9.0
Steam Generation	$14.4
Ash Handling	$11.4
Pressurization Gas	$6.0
Utilities/Offsites	
Water Treatment Systems	$6.6
Cooling Water Systems	$6.1
Fuel Oil System	$0.6
Inst./Plant Air	$2.0
Flue Gas Handling	$10.9
Chimney	$1.7
Fire Protection	$1.9
Steam Turbine	$24.7
Electrical Distribution	$4.2
Buildings	$31.4
Rail Spur	$0.0
Infrastructure	$3.0
TOTAL DIRECT COST	$248.8
Field Management (11.5%)	$28.6
TOTAL FIELD COST	$277.4
Home Office Fee and Services (11.5%)	$28.6
SUB TOTAL	$306.0
Contingency (14%)	$42.8
TOTAL PLANT COST	$348.8
Taxes/Duties/Owner's Cost/Initial Chemicals (12% of TPC)	$41.9
SUB TOTAL	$390.7
Interest During Construction (24.5%)	$95.7
TOTAL CAPITAL REQUIREMENT	$486.4 $1,713 $/kW

The costs shown have been scaled and inflated from the BC data unless otherwise indicated. The coal receiving, storage and reclaim facilities are available due to the development for the Point Aconi 165 MWe CFBC. The sorbent reception and storage is also available due to the development at Point Aconi.

The sand handling, combustor feed, feed handling, gasification, hot gas cleaning, gasification air compressor and gas turbine have all been adjusted to account for scale according to the equation given above.

The CFB combustor cost was scaled from the NSPC Point Aconi data.

The cooling water costs are at 20% of the BC amounts. This reflects the previous installation of the cooling water off-shore facilities, pumphouse, cooling water system intake and discharge and the outfall structure.

Electrical distribution costs are 20% of the BC costs due to previous development.

The rail spur costs were waived as they have been developed for the 165 MWe CFBC.

The infrastructure costs are at 20% of the BC values due to previous development.

A contingency of 14% has been added to reflect the lack of accuracy of the capital cost estimates.

Start-up Costs and Working Capital

Start-up costs have been estimated at one month's operation cost (excluding coal) + 25% of one month's fuel cost + 2% of the power plant capital cost.

Working capital has been determined to be equal to the cost of 60 days of coal, water, limestone, ash and spent sorbent disposal plus 1% of the capital requirement.

5.2.2 Operating and Maintenance Costs

Table 5.3 shows the fixed and variable operating costs that have been developed from the BC data. The fixed operating costs have not been scaled from the 350 MWe BC costs. Linear scaling has been applied to the variable operating costs. Solid waste disposal costs have been calculated based on the coal feed rate and analysis.

Table 5.3 also shows the operating data used in the cost analysis.

Table 5.3

Operating Data for the Conceptual VEW System Located at Point Aconi

Coal Related Operational Data	
HHV	31 MJ/kg (in situ)
Sulphur	4.0%
Ash	7.0%
Price	$65/tonne ($2.10/GJ) (as per Table 3.3)
Average Yearly Use	525,292 t/y
Cost (daily)	$93,545/d
Ash Related Operational Data	
Ash Generation Rate	100.7 t/d
Ash Disposal Unit Cost	$5.5/t
Ash Disposal Cost	$554/day
Limestone and Spent Sorbent Operation Data	
Limestone Unit Cost	$15/t
Limestone Utilization Rate	445t/d
Limestone Cost	$6,675/d
Spent Sorbent Disposal Cost	$5.5/t
Spent Sorbent Generation Rate	605t/d
Sorbent Disposal Cost	$3,329/t
Fixed Operating Costs	
Operating Labour	$4.3
Maintenance Labour	$1.7
Administrative and Support	$1.1
Total Fixed Costs	$7.2
Variable Operating Costs	
Maintenance Materials	$6.8
Contract Labour	$2.4
Sorbent, Chemicals & Catalysts	$4.1
Solid Waste Disposal	$1.4
Make-up Water	$0.4
Total Variable Costs	$15.1
TOTAL OPERATING AND MAINTENANCE	$22.2

5.2.3 Cost of Electricity Sensitivity Analysis

The levelized cost of electricity (COE) has been determined by using a discounted cash flow (DCF). The various levelized costs have been determined as follows.

The capital recovery factor (a_n) is given by:

$$\frac{1}{a_n} = \frac{i \cdot (1 + i)^n}{(1 + i)^n - 1}$$

where i is the interest rate and n is the period in years.

The present value factor (V_n) is given by:

$$V_n = \frac{1}{(1 + i)^n}$$

The levelized capital charge is determined using:

$$\textit{Captial Charge} = \frac{\dfrac{CR}{a_n} + \dfrac{WC \cdot (1 - V_n)}{a_n} \; x\ 1000}{EP} \quad (\textit{mils/kWh})$$

where CR is the capital required, WC is the working capital and EP is the average annual power production.

The levelized operating and maintenance costs are given by:

$$\textit{Operating} + \textit{Maintainence Costs} = \frac{OMC}{EP} \; x\ 100 \quad (\textit{mils/kWh})$$

where OMC is the operating and maintenance costs.

The levelized fuel costs are given by:

$$Fuel\ Costs = \frac{FC}{EP} \times 100 \quad (mils/kWh)$$

where FC is the annual fuel cost.

The levelized cost of electricity (COE) is equal to the sum of the levelized capital charge + levelized operating & maintenance costs + the levelized fuel costs.

The data presented above were used in the equations given to arrive at the values shown in Table 5.4.

Table 5.4

Levelized Cost Factors Used to Determine the Cost of Electricity for the Conceptual VEW System Located at Point Aconi

Values used to Determine the Cost of Electricity	
Power and Gasification Plant Cost	$1713/kWe
Start-up Costs	$38/kWe
CR (capital required)	$1751/kWe ($497M)
WC (working capital)	$40/kWe ($11.2M)
OMC (operating & maintenance costs)	$22M
FC (annual fuel costs)	$34.1M
EP (average annual power (net) production)	1990 GWh/y

The base case COE was determined using an 8% DCF rate of return. The discounted rate of return on investment was varied from 5 to 12% to show the effect on the COE. This effect is shown in Figure 5.1.

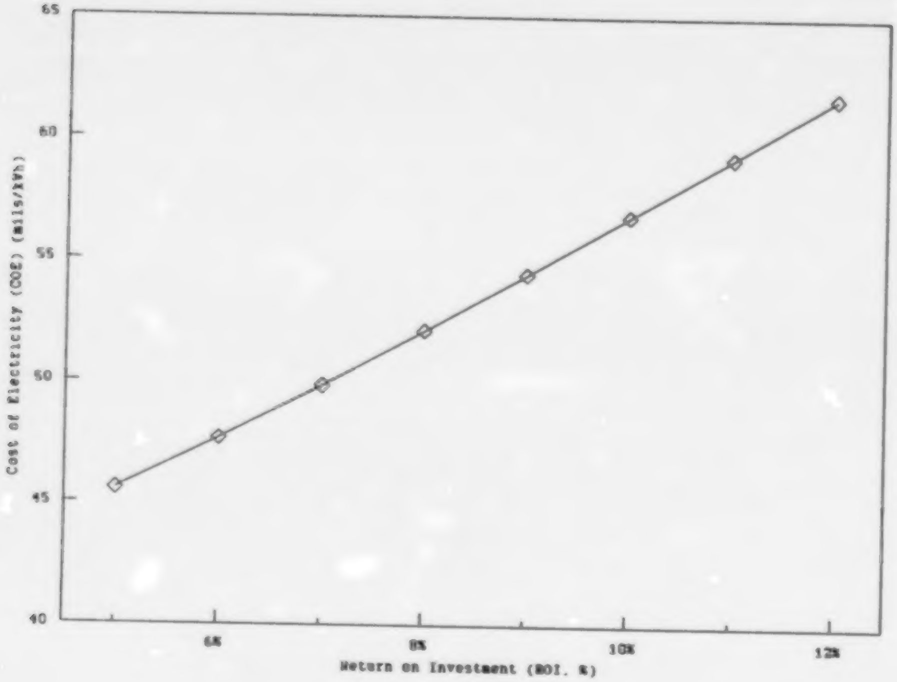


Figure 5.1

Sensitivity of Cost of Electricity to the Rate of Return on Investment for the VEW Conceptual Partial Gasification Co-Product Process

The base case capital costs were varied by ± 40% to evaluate the COE sensitivity. The results are shown in Figure 5.2.

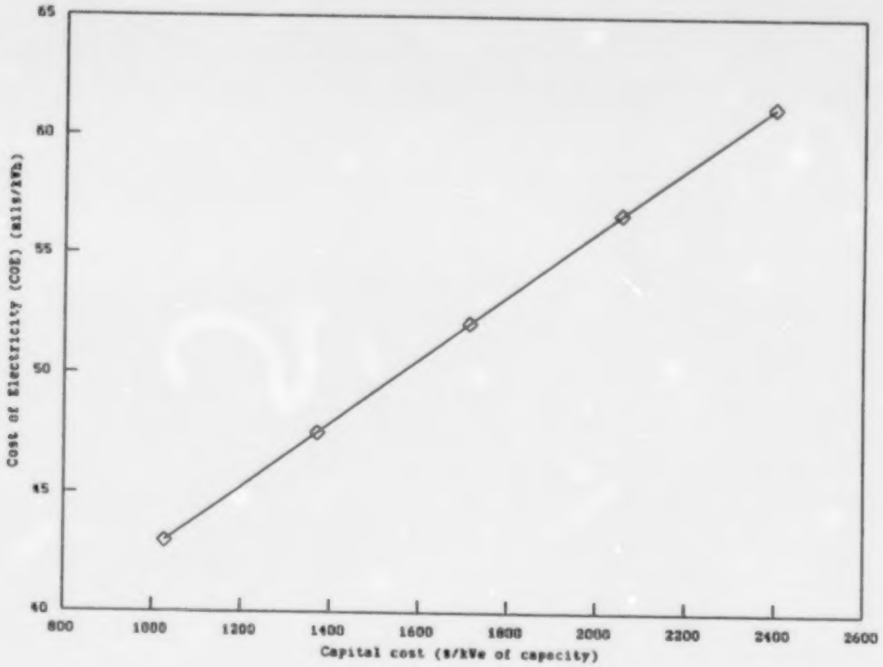


Figure 5.2

Sensitivity of Cost of Electricity to the Capital Cost for the VEW Conceptual Partial Gasification Co-Product Process

6 INSTITUTE OF GAS TECHNOLOGY (IGT) MILD GASIFICATION CONCEPTUAL PROCESS; PRODUCT UTILIZATION, PROCESS DESCRIPTION AND ECONOMICS

This section of the report describes the potential utilizations for the products and the economic analysis for the Institute of Gas Technology Co-Product Process.

6.1 Conceptual Process Scheme

Based on the market survey for the char and condensables, three char utilization options for integrating the mild gasification process with an existing coal preparation plant were chosen as shown below:

1) Char/Coal Blending for Utility Fuel

2) Metallurgical Form Coke Production

3) Smokeless Domestic Fuel Production

Figures 6.1 to 6.3 show the conceptual process flow diagrams for the three schemes noted above. In each figure, the approximate mass flow rates for each stream are shown in tons/day. A description of the process steps is given below.

All three schemes utilize the same condensable co-product recovery concepts used for conventional recovery-type coal coking systems. A more optimum process may be proposed as advances are made in the shale and coal pyrolysis recovery trains. This may affect the distribution of the liquid products but not the overall quantity.

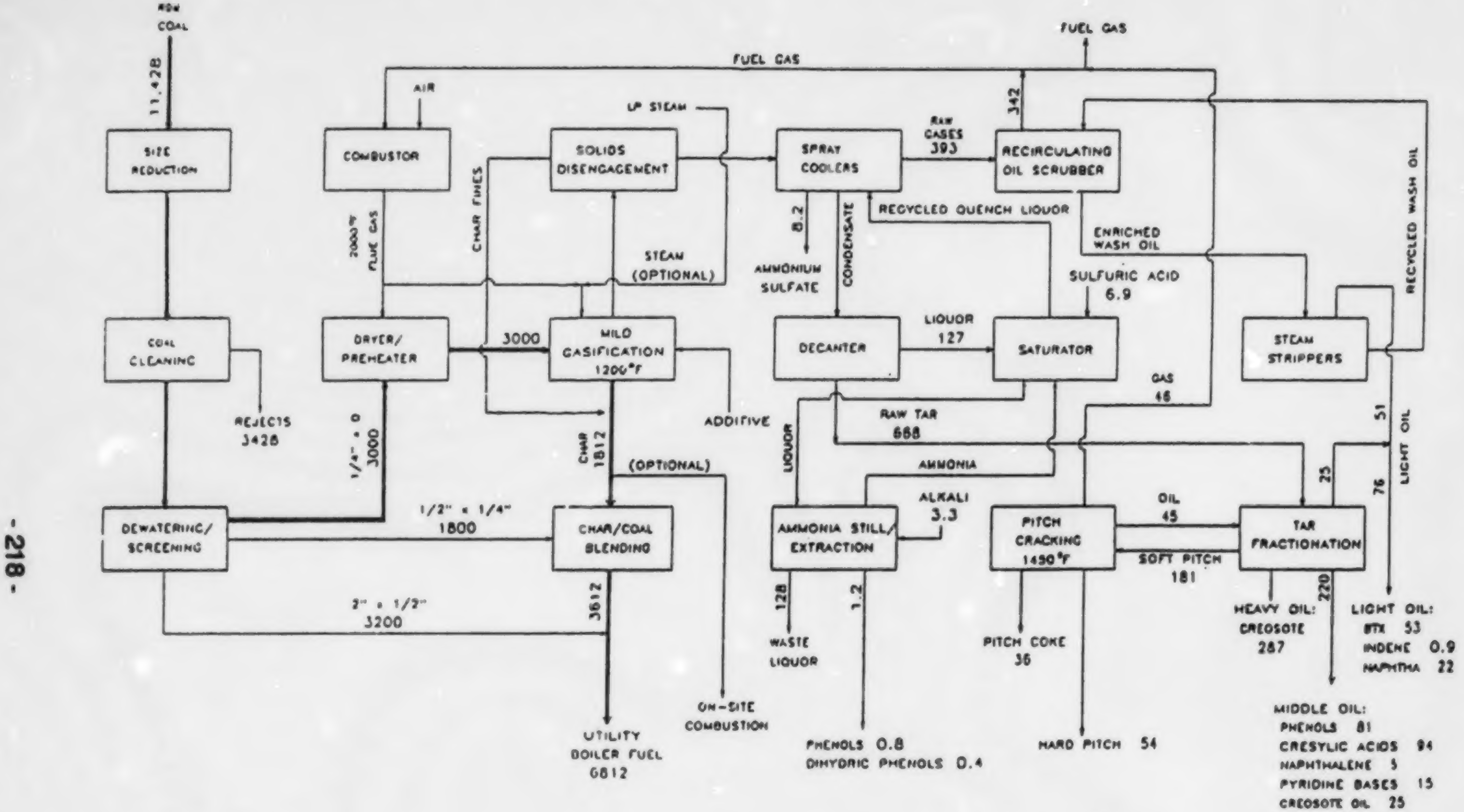

RAW COAL
11,428
SIZE REDUCTION
COAL CLEANING
REJECTS 3428
DEWATERING/ SCREENING
1/4" x 0 3000
1/2" x 1/4" 1800
2" x 1/2" 3200
FUEL GAS
AIR
COMBUSTOR
2000°F FLUE GAS
DRYER/ PREHEATER
CHAR FINES
3000
LP STEAM
SOLIDS DISENGAGEMENT
STEAM (OPTIONAL)
MILD GASIFICATION 1200°F
ADDITIVE
CHAR 1812
(OPTIONAL)
CHAR/COAL BLENDING
3612
UTILITY BOILER FUEL 6812
ON-SITE COMBUSTION
SPRAY COOLERS
RAW GASES 393
8.2
AMMONIUM SULFATE
CONDENSATE
RECYCLED QUENCH LIQUOR
DECANTER
LIQUOR 127
RAW TAR 688
LIQUOR
AMMONIA
AMMONIA STILL/ EXTRACTION
ALKALI 3.3
128
WASTE LIQUOR
1.2
PHENOLS 0.8
DIHYDRIC PHENOLS 0.4
FUEL GAS
342
RECIRCULATING OIL SCRUBBER
ENRICHED WASH OIL
SULFURIC ACID 6.9
SATURATOR
GAS 46
PITCH CRACKING 1450°F
PITCH COKE 36
HARD PITCH 54
OIL 45
SOFT PITCH 181
STEAM STRIPPERS
RECYCLED WASH OIL
51
LIGHT OIL
25
76
TAR FRACTIONATION
220
HEAVY OIL: CREOSOTE 287
LIGHT OIL:
BTX 53
INDENE 0.9
NAPHTHA 22
MIDDLE OIL:
PHENOLS 81
CRESYLIC ACIDS 94
NAPHTHALENE 5
PYRIDINE BASES 15
CREOSOTE OIL 25

Figure 6.1

Option I: IGT Conceptual Process Flow Diagram, Char/Coal Blending for Utility Fuel Production and Tar Fractionation
Wooten et al., 1988
(mass flows are in tons/day)

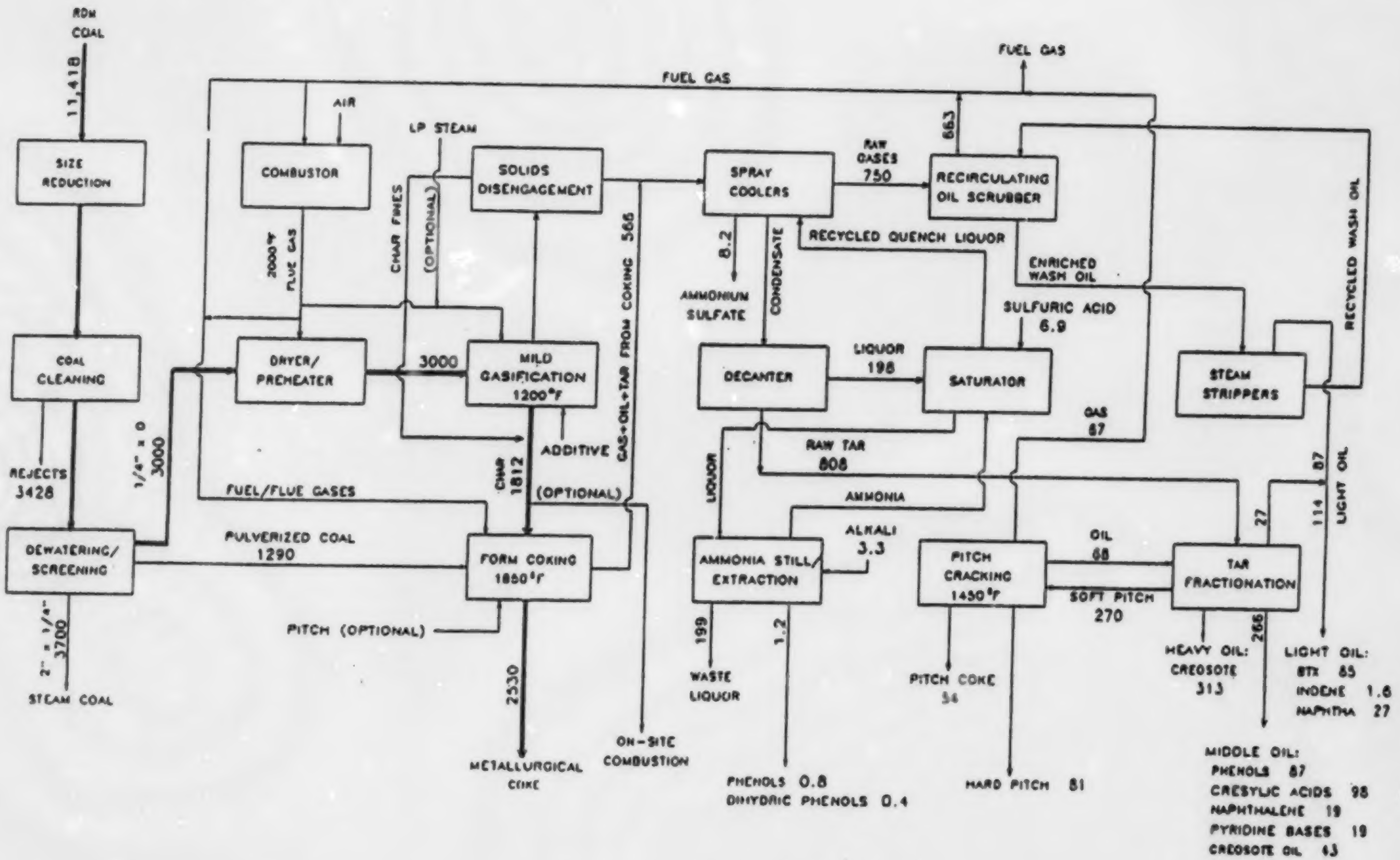

RAW COAL
11,418
SIZE REDUCTION
COAL CLEANING
REJECTS 3428
DEWATERING/ SCREENING
2" x 1/4" 3700
STEAM COAL
1/4" x 0 3000
COMBUSTOR
AIR
2000°F FLUE GAS
DRYER/ PREHEATER
3000
MILD GASIFICATION 1200°F
ADDITIVE
CHAR 1812
CHAR FINES
(OPTIONAL)
LP STEAM
SOLIDS DISENGAGEMENT
FUEL/FLUE GASES
PULVERIZED COAL 1290
FORM COKING 1850°F
PITCH (OPTIONAL)
2530
METALLURGICAL COKE
(OPTIONAL)
GAS+OIL+TAR FROM COKING 566
ON-SITE COMBUSTION
FUEL GAS
FUEL GAS
SPRAY COOLERS
RAW GASES 750
RECIRCULATING OIL SCRUBBER
663
8.2
AMMONIUM SULFATE
CONDENSATE
RECYCLED QUENCH LIQUOR
ENRICHED WASH OIL
SULFURIC ACID 6.9
DECANTER
LIQUOR 198
SATURATOR
RAW TAR 808
LIQUOR
AMMONIA
ALKALI 3.3
AMMONIA STILL/ EXTRACTION
199
WASTE LIQUOR
1.2
PHENOLS 0.8
DIHYDRIC PHENOLS 0.4
GAS 87
PITCH CRACKING 1450°F
PITCH COKE 54
HARD PITCH 81
OIL 68
SOFT PITCH 270
TAR FRACTIONATION
STEAM STRIPPERS
RECYCLED WASH OIL
27
114
87
LIGHT OIL
266
HEAVY OIL: CREOSOTE 313
LIGHT OIL:
BTX 85
INDENE 1.6
NAPHTHA 27
MIDDLE OIL:
PHENOLS 87
CRESYLIC ACIDS 98
NAPHTHALENE 19
PYRIDINE BASES 19
CREOSOTE OIL 43

Figure 6.2

Option II: IGT Conceptual Process Flow Diagram, Metallurgical Form Coke Production and Tar Fractionation
Wooten et al., 1988
(mass flows are in tons/day)

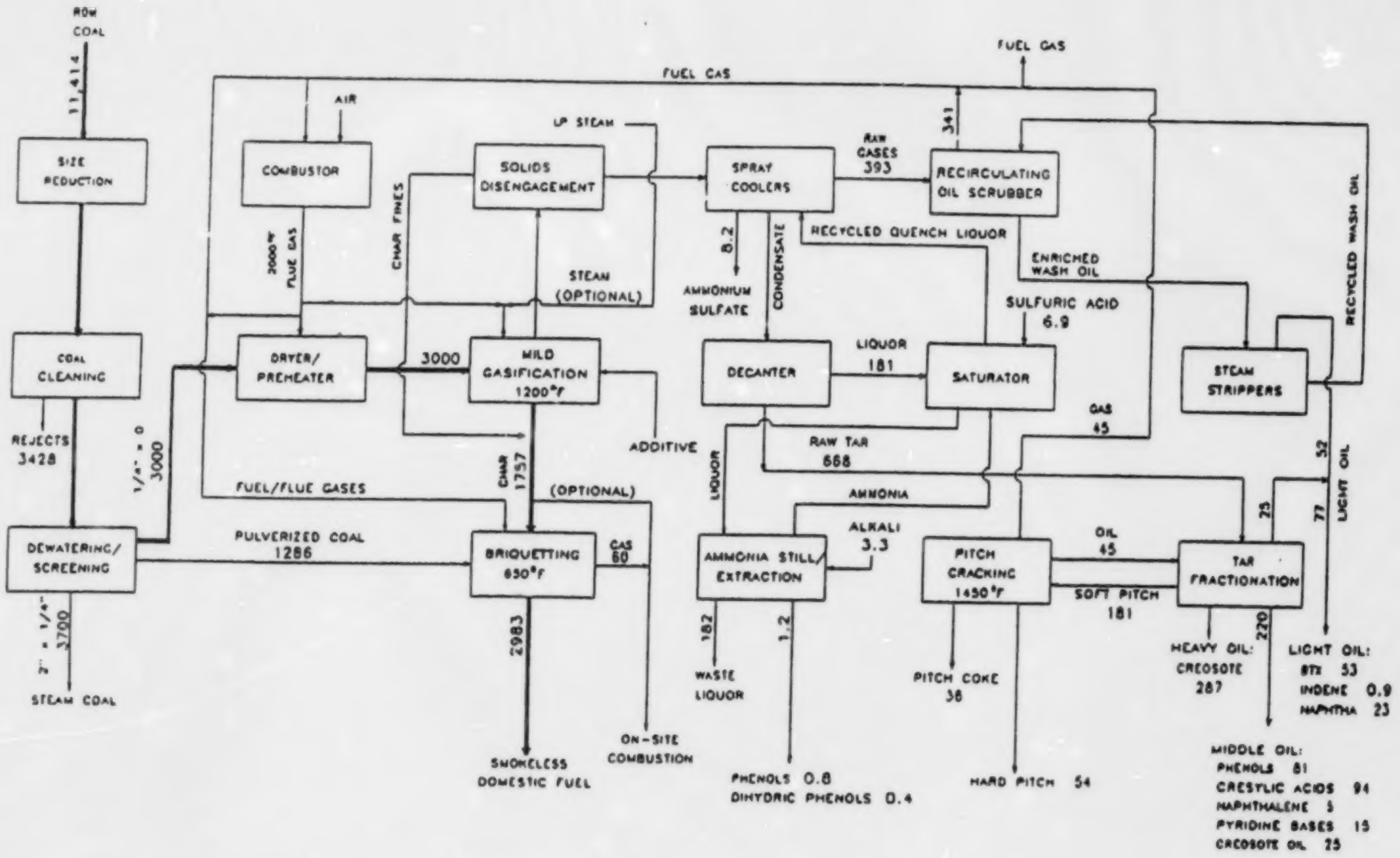

RAW COAL
11,414
SIZE REDUCTION
COAL CLEANING
REJECTS 3428
DEWATERING/ SCREENING
1/4" x 0
3000
2" x 1/4"
3700
STEAM COAL
FUEL GAS
AIR
COMBUSTOR
2000°F FLUE GAS
DRYER/ PREHEATER
CHAR FINES
FUEL/FLUE GASES
PULVERIZED COAL
1286
LP STEAM
SOLIDS DISENGAGEMENT
STEAM (OPTIONAL)
3000
MILD GASIFICATION 1200°F
ADDITIVE
CHAR 1757
(OPTIONAL)
BRIQUETTING 650°F
GAS 60
2983
SMOKELESS DOMESTIC FUEL
ON-SITE COMBUSTION
SPRAY COOLERS
RAW GASES 393
8.2
AMMONIUM SULFATE
CONDENSATE
RECYCLED QUENCH LIQUOR
DECANTER
LIQUOR 181
RAW TAR 668
LIQUOR
AMMONIA
AMMONIA STILL/ EXTRACTION
ALKALI 3.3
182
WASTE LIQUOR
1.2
PHENOLS 0.8
DIHYDRIC PHENOLS 0.4
FUEL GAS
341
RECIRCULATING OIL SCRUBBER
ENRICHED WASH OIL
SULFURIC ACID 6.9
SATURATOR
GAS 45
PITCH CRACKING 1450°F
PITCH COKE 38
HARD PITCH 54
OIL 45
SOFT PITCH 181
TAR FRACTIONATION
HEAVY OIL: CREOSOTE 287
220
MIDDLE OIL:
PHENOLS 81
CRESYLIC ACIDS 94
NAPHTHALENE 5
PYRIDINE BASES 15
CREOSOTE OIL 25
RECYCLED WASH OIL
STEAM STRIPPERS
52
25
77
LIGHT OIL
LIGHT OIL:
BTX 53
INDENE 0.9
NAPHTHA 23

Figure 6.3

Option III: IGT Conceptual Process Flow Diagram, Smokeless Fuel Production and Tar Fractionation
Wooten et al., 1988
(mass flows are in tons/day)

6.1.1 Char/Coal Blending for Utility Fuel Process

The production rate of the various products from Option I are shown in Table 6.1. These rates are based on a total feed rate of 2722 t/d dry coal to three gasifier reactors.

Table 6.1

IGT Mild Gasification Co-Product Conceptual Process, Option I, Char/Coal Blending for Utility Fuel Production and Tar Fractionation, Annual Production Rates

Product	Annual Production
	t/yr
Pitch (roofing)	16,093
Pitch Coke (liquid)	10,728
Light Oil	22,649
Medium Oil	65,562
Creosote Oil	85,529
Phenols	358
Ammonium Sulphate	2,444
Char/Coal Blend	1,030,045

The following description of the processes are taken directly from Wooten et al., 1988.

"The conceptual mild gasification process consists of 16 steps, each of which is comprised of several unit operations and unit processes. The stream flow rates are based on a dry basis coal fed to three mild gasification reactors of equal capacity totalling 2722 t/d. All input and output stream flows are given in ton/d on a dry coal basis. The principal process steps are:

Size Reduction

"Run-of-mine coal (10,367 t/d) is crushed and screened to yield a 15 cm x 0 cm raw coal feed. Based on Peabody preparation plant experience this product contains up to 37.5% under 0.63 cm which is considered an undersize fraction. The entirety of the sized coal is sent to the coal cleaning step.

Coal Cleaning

"The raw coal feed from the size reduction step is subjected to conventional coal washing to reduce ash and sulphur content. During washing the coal is further split into a coarse (15 cm x 2.5 cm), middling (2.5 cm x 1 mm), and fine (1 mm x 100 mesh) streams which are each treated to a different washing scheme. During washing, coarse refuse (gob) and the -100 mesh stream are rejected and are expected to total about 3110 t/d.

Dewatering/Screening

"The coarse coal fraction is screened and crushed to a top size of 5.1 cm. The middling and fine streams are fed to mechanical dewatering devices where surface and interstitial water are removed. At this point, a screening operation would be incorporated into the process that would yield a coarse product (5.1 cm x 1.3 cm), a middling product (1.3 cm x 0.63 cm), and a fine product (0.63 cm x 0 cm) stream. The fine product stream, approximately 2722 t/d would be fed to the dryer/preheater as preparation for mild gasification. The middling fraction, amounting to about 1633 t/d, would be sent to the char/coal blending step. The char/coal blend will be further blended with the coarse product, approximately 2903 t/d producing a total of 6812 t/d of char/coal blend utility fuel.

Dryer/Preheater

"The 0.63 cm x 0 cm cut from the dewater/screening step is dried to about 5% to 10% moisture and preheated to approximately 232°C by exposure to hot flue gases from the in-plant combustor, which produces heat from the combustion of product fuel gases (or alternatively char fines). The 2722 t/d (dry basis) output from this step is delivered directly to the mild gasification reactors.

Mild Gasification

"The mild gasification step consists of three gasifiers, each with 907 t/d capacity. The mild gasifiers operate at a temperature of 649°C and a pressure between atmospheric and 343 KPa (gauge). The fluidizing/entraining gas is a mixture of hot flue gases from the fuel gas combustor and (optionally) steam produced from waste heat recovery. The coal feed may include an optional solid phase sulphur-capture additive.

"Two output streams exit the mild gasification step: the solids- and tar-laden gas; and the devolatilized char mixed with the partially sulphated additive, if that option has been utilized.

Solids Disengagement

"The solids-laden off-gas from the mild gasification step exits the top of the reactors, and char fines are removed by one or more high-temperature cyclones. The char fines are then blended back with the bottom char product. Alternatively they may be recycled to the retort to provide a portion of the required heat, or may be burned in place of fuel gas to provide the hot recycle gas stream.

Char Blending

"The hot (649°C) coal char exiting the mild gasification operation, totalling 1644 t/d after char fines recovery, is blended with the 1.3 cm x 0.63 cm middling cut from the preparation plant dewater/screening step. The sensible heat contained in the char is used to dry the coal fines; also, the blending of hot char and cold coal fines is performed in a controlled manner in order to increase the average particle size by the agglomeration of coal particles as the temperatures approach the plastic range. This should improve the shipping and handling characteristics of the resulting utility fuel product. The partially agglomerated char/coal mixture is then blended with the 5.1 cm x 1.3 cm fraction from the dewatering step to produce a solid fuel product with the desired properties.

"As an option, part of the hot char is removed prior to the blending step for in-plant combustion to provide process heat.

"If a solid additive such as lime has been used in the mild gasification step, the additive could remain with the char in order to provide the solid fuel mixture with sulphur-capture properties during char combustion in a subsequent step.

Spray Coolers

"Gas cooling and recovery of condensables are achieved in this step by standard coal-tar quenching and recovery methods, as described previously. For simplicity, several operations are categorized together with the spray cooling process step. Recycled acidic quench liquor, saturated in ammonium sulphate and containing a 5% excess of sulphuric acid, is used in one or more spray quench coolers. The volatile product stream is ultimately cooled to about ambient temperature in this operation. The raw gas exiting this step, which is saturated at about 21°C with light oil, amounts to 357 t/d.

"A portion of the naphthalene produced in the mild gasifier will be condensed from the product gases in the final cooling state, where it is skimmed from the aqueous layer or extracted with an oil wash.

"Ammonium sulphate in crystalline form is recovered from the quench liquor in a crystallizer; this is made possible by the super-saturation of the quench liquor with ammonia from the product gases.

"The condensate, consisting of an organic condensable phase and an aqueous liquor, is delivered to a decanting step for separation.

Decanter

"The organic and aqueous phases of the condensate are separated by decanting, yielding 606 t/d of raw tar and 115 t/d of liquor. The aqueous liquor, containing ammonia, phenols, and other water-soluble organic compounds, is sent to the saturator which supplies quench liquor to the spray coolers. The organic condensables are sent to the tar distillation and fractionation step.

Saturator

"In this process step, ammonia recovered from the decanted liquor by distillation is recombined with the incoming liquor and sulphuric acid to form the acidic saturated ammonium sulphate solution used as a quench liquor.

Ammonia Still/Extraction

"The excess liquor from the saturator is treated with alkali or lime to release ammonia, which is steam-stripped from the liquor and recycled to the saturator. The waste liquor is then cooled and extracted with aromatic solvents to recover 1.1 t/d of phenols, including approximately 0.4 t/d of dihydric phenols (catechol, resorcinol and their derivatives), which are then regenerated or "sprung" from the solvents by treatment with aqueous alkali. The waste liquor stream, exclusive of any added water from steam stripping or other auxiliary inputs, amounts to 114 t/d.

Recirculating Oil Scrubber

"Raw non-condensable gases exiting the condensation step are contacted with a suitable solvent which can be a fraction of the middle oil recovered from fractionation of the coal tar, or a petroleum-derived solvent. Light oils, including benzene, toluene, and xylenes (BTX) are absorbed from the product gas by this method. The "enriched" wash oil is sent to steam stripping to recover the light oils.

Steam Strippers

"Steam distillation is used to remove the light oils from the wash oil. The light oil consists of approximately 74 t/d of BTX, plus small amounts of indene and coal tar naphtha. The recovered light oils are combined with light oils fractionated from the condensable liquids in the tar fractionation step, and the lean wash oil is recycled back to the recirculating oil scrubber.

Tar Fractionation

"In order to recover the high-value products from the raw organic condensables produced by mild gasification, a fractionation system is required. This process step involves primary distillation and steam stripping of light oils. The removal of tar acids by alkali treatment, recovery of tar bases by treatment with acid, and further fractionation of the major fractions to recover BTX, indene, tar naphtha, light phenols, cresylic acids, naphthalene, pyridine bases and creosote oil cuts are options.

"In primary distillation, the condensable product, following filtration, is flashed-distilled to produce four fractions:

- Light oil, distilling up to 180°C, approximately 23 t/d
- Middle oil, distilling up to 310°C, approximately 200 t/d
- Heavy oil, distilling up to 400°C, approximately 260 t/d
- Soft pitch, distillation residue, approximately 164 t/d.

"The input streams to the tar fractionation step are the raw organic condensable phase from the decanter and the oil recovered from the pitch cracking operation.

Pitch Cracking

"In order to produce a hard pitch of premium value, suitable for electrode binder use, it is necessary to further process the soft pitch produced by the low-temperature mild gasification. A simple thermal treatment of 788°C is used to crack the pitch to produce 33 t/d of pitch coke suitable for use as graphite, 49 t/d of hard pitch, 41 t/d of oil which is recycled back to the tar fractionation step, and 41 t/d medium-Btu fuel gas which is combined with the primary fuel gas recovered from the oil scrubber step.

Option II. Metallurgical Form Coke Production

"In this option, char and pulverized coal are combined to manufacture metallurgical form coke. The coarse fraction from the dewatering/screening step in the coal preparation circuit is used for sale as a utility fuel.

"All of the process steps except the form coking operation are similar to those described under Option I. The addition of a volatile stream from the form coke high-temperature carbonization step changes the yield and distribution of gaseous and condensable products from the overall process. This is reflected in the t/d yields of the product streams as shown in Figure 6.2.

The annual production rates of the products from Option II are shown in Table 6.2.

Table 6.2

IGT Mild Gasification Co-Product Conceptual Process, Option II, Form Coke Production and Tar Fractionation, Annual Production Rates

Product	Annual Production
	t/yr
Pitch (roofing)	24,139
Pitch Coke (liquid)	16,093
Light Oil	33,973
Medium Oil	67,350
Creosote Oil	93,277
Phenols	358
Ammonium Sulphate	2,444
Form Coke	753,965

Form Coking

"In this step, the hot (649°C) char from the mild gasifier and the solids disengagement step are combined and blended with cleaned, dewatered pulverized bituminous coal in a hot briquetting operation. In this fashion, the plastic properties of the raw coal are utilized to provide the binding strength of the briquette, with the heat being supplied by the sensible heat of the char, supplemented by external heat from combustion of a portion of the char, if necessary. As an option, pitch from the tar fractionation or pitch cracking step can be added as an additional binder, if required to improve the physical properties of the briquette. The briquettes, prior to carbonization, will contain 20% to 25% volatile matter, based on a feed of Illinois No. 6 coal.

"The briquettes are then carbonized at 1010°C in either a vertical shaft furnace or another type of suitable furnace to produce form coke with 3% to 5% volatile matter. The caking characteristics of the coal fines should contribute to the production of briquettes with the required physical properties. The off-gas from the high-temperature carbonizer is combined with the gas stream exiting the mild gasification solids disengagement step, for additional recovery of fuel gas and condensables in the downstream processing steps.

Option III. Smokeless Domestic Fuel Production

"In this option, shown in Figure 6.3, the mild gasification char is combined with pulverized coal to make a briquette as in Option II. The briquettes are treated to produce a smokeless domestic fuel briquette suitable for use in home stoves or for district heating purposes.

"Except for the briquetting/curing step, all of the process steps are similar to those described in Option I. The yields of gaseous and condensable products from the mild gasifier are identical to those shown for Option I.

The annual production rates for Option III are shown in Table 6.3.

Table 6.3

IGT Mild Gasification Co-Product Conceptual Process, Option III Smokeless Fuel and Tar Fractionation, Annual Production Rates

Product	Annual Production
	t/yr
Pitch (roofing)	16,093
Pitch Coke (liquid)	7,748
Light Oil	22,947
Medium Oil	65,562
Creosote Oil	85,529
Phenols	358
Ammonium Sulphate	2,444
Domestic Fuel	888,964

Briquetting

"The combined hot char and hot char fines from the mild gasification and solids disengagement steps are combined with pulverized coal, making use of the sensible heat contained in the char to achieve a controlled plasticity in the char/coal mixture. This provides the binding mechanism for briquette production. The briquettes are then cured by holding at a temperature of 343°C, resulting in the production of about 54 t/d of fuel gas. This fuel gas is used to provide additional heat for in-plant uses, such as for the briquette curing operation.

"Optionally, soft or hard pitch can be used as an auxiliary binder if it is deemed necessary to improve the physical properties of the briquettes.

"The briquettes will contain approximately 20% volatile matter, based on a feed of Illinois No. 6 coal."

6.2 Product Utilization Options

The IGT mild gasification process produces solids, condensable hydrocarbons and gas from the coal. This section will describe the products and how they might be utilized.

6.2.1 Solid Co-Products Utilization

Three uses for the char have been identified: char/coal blending for utility fuel, metallurgical form coke and smokeless domestic fuel. The following sections describe the products; relates their characteristics to those needed for each application and briefly discusses the market for the products.

6.2.1.1 Char/Coal Utility Fuel

The char from the gasification process could be blended with thermal coal to produce a utility fuel. The local market for a blended thermal fuel of char and coal would be the Nova Scotia Power Corporation (NSPC). The source of coal and the location and magnitude of NSPC current consumption is shown in Table 6.4.

Table 6.4

Nova Scotia Power Corporation Coal Source and Consumption

NSPC Facility Information		Coal Source		
Plant Name	Average Capacity Factor (%)	Mine/Seam	Capacity Net (MW_e)	Coal Utilization (1000's t/y)
Lingan	75	Prince/Hub & Lingan/Harbour[a]	642	1,460
Glace Bay	19	Prince/Hub & Lingan/Harbour[a]	105	87
Point Tupper II	70	Prince/Hub & Lingan/Harbour[a]	149	354
Trenton V	57	Westray/Foord[b]	149	280
Trenton VI	70[c]	Westray/Foord[d] (regular, washed)	160	400
Point Aconi	70[c]	Prince[e]	165	450

Notes for Table 6.4.

- a - Prince run of mine coal is blended with Lingan and Phalen middlings from the Victoria Junction wash plant to produce a thermal coal with 3% sulphur. Typically the blend is 50% Prince and 50% VJ middlings.
- b - Trenton V is in the process of switching from the coal described in "a" above to Westray premium washed coal.
- c - Not operational
- d - Trenton VI will use Westray regular grade coal.
- e - Point Aconi will start on stream in 1993. At that time it will start to use Prince coal leaving a smaller fraction of the Prince coal to be blended with the Lingan and Phalen thermal streams. This will result in a reduction in the sulphur content of the coal going to Lingan, Tupper II and Glace Bay.

In addition, small coal companies may supply 10,000 to 20,000 t/y coal to Glace Bay and some of the other plants.

Based on the information in Table 6.4, the total consumption of NSPC will be approximately 2,981,000 t/y, with approximately 2,950,000 t/y coming from the Cape Breton coal fields after the conversion to the Westray coal at Trenton is completed. The char production rate from the char/coal blend option (from the data in Figure 6.1) is 539,994 t/yr which could easily be utilized even with reductions in coal utilization at some stations after Aconi comes on line.

The analysis of the chars produced from the IGT Mild Gasification process are presented in Table 6.5 as calculated from the data presented in Tables 3.50, 3.51 and 3.53 to 3.57. The values shown for tests MG-28 and MG-30 are for the fluid bed char and that for tests IST-10 to 13 are for the char plus the diluent. This data has been generated as a result of the gasification of the West Virginia coal whose analysis as shown in Table 3.45. This analysis should be compared to that of the Nova Scotia coals shown in Tables 3.3 to 3.6 and 3.7 to 3.10. Approximations of the solid product analysis will be made with consideration of the variation in the coal characteristics.

Table 6.5

Analysis of Char Produced From the IGT Mild Gasification Process

	MG-28 Char Mix	MG-30 Char Mix	IST-10 Char & Diluent	IST-11 Char & Diluent	IST-12 Char & Diluent	IST-13 Char & Diluent
Carbon (%)	83.51	82.69	83.03	82.28	83.67	83.37
Hydrogen (%)	0.63	0.82	1.63	1.93	1.03	1.55
Oxygen (by dif.) (%)	0.78	2.17	2.45	2.50	1.36	2.39
Nitrogen (%)	1.25	1.38	1.39	1.62	1.34	1.55
Sulphur (%)	0.94	0.96	1.60	1.16	0.71	0.84
Ash (%)	12.78	11.98	9.89	10.51	11.89	10.31

The fixed carbon (FC) values in the WV and CB NS coals do not differ significantly, therefore under similar reactor conditions, a value of approximately 82% FC would be expected for the char from CB NS coals (see Table 3.47 for WV char data).

IGT has presented some data for undiluted char material in which the ratio of the ash in the char to that in the parent coal is given to be approximately 1.6. This ratio would suggest that the Cape Breton coals would produce a char with ash contents between 8 and 13% and the Westray coal would produce a char with nearly 35% for the regular coal grade.

Volatile matter ratios in the char to that in the parent coal were in the 0.24 to 0.42 range which leads to a prediction of volatile matter in char from CB coal between 9 and 16% and between 7 and 12% for the Westray coal.

Sulphur ratios ranged from 0.58 to 1.19 (char to coal) for the IGT WV data. The ratio of organic to pyritic sulphur in the WV and CB coals are 2/1 and 1/2 respectively. The majority of the organic sulphur is released during gasification while the majority of the pyritic sulphur remains with the char. Therefore, this variation in the sulphur form will have a significant effect on the percentage of the parent coal sulphur that remains in the char. Assuming 100% of the pyritic sulphur in the parent coal remained in the char and partial gasification yields 60% char, the total percentage sulphur in the char would be as shown in Table 6.6.

Table 6.6

Projected Sulphur in Chars Generated From the Mild Gasification of Nova Scotia Coals in the IGT Mild Gasification Process

Coal	% Sulphur
Prince	4.67
Lingan	2.33
Phalen	2.83
Westray	0.33

HHV's in the 29 MJ/kg range could be expected. This will be a function of the parent coal characteristics and the reactor conditions.

The following analysis will be in consideration of an IGT partial gasification plant operating on Prince coal. This is done in light of the proximity of the Prince mine to the NSPC Point Aconi 165 MWe CFBC facility that could accept the char as a fuel.

The char characteristics that would be expected from the mild gasification of Prince coal is shown in Table 6.7.

Table 6.7

Projected Analysis of Char Produced From the Mild Gasification of Prince N.S. Coal in the IGT Process

PROXIMATE	
Volatile Matter	9 - 16%
Fixed Carbon	82%
Ash	8 - 13%
ULTIMATE	
C	82%
H	1.5%
N	1.5%
S	4%
Ash	10%
O	1.0%
HHV	29 MJ/kg

The hypothetical analysis presented in Table 6.7 would result in a sulphur release rate of 1.38 g/MJ as a result of the combustion of the char. This compares to 1.27 g of sulphur/MJ for the Prince coal. This represents approximately a 9% increase in the sulphur emission rate. This figure must be viewed as a very rough approximation at best.

The feasibility of using the char or a char/coal blend as a substitute boiler fuel will be somewhat dependant upon the performance of the char in the combustor. If the char were to be fed to the Point Aconi CFBC, the characteristics of the char would be far less critical than if the utilization were to be in a conventional pulverized fired coal type boiler.

In the CFBC application, the sulphur can be fixed using a sorbent such as limestone, the higher ash content can be accommodated, variations in the ash fusion characteristics are less important due to the operation of the combustor at 850°C and the char particle size distribution may be acceptable as is from the gasifier system.

Utilization of the char in a conventional pulverized coal fired design unit would necessitate more rigid evaluation of the char's combustion performance. Characteristics such as: moisture, FC, VM, ash, sulphur, HHV, ash fusion temperature and size distribution would need to be examined in relation to the end use conditions. Some difficulty may arise in making evaluations of the char when comparing its properties to that of coal. The reactivity of char in a combustion environment is to a large extent dependant upon the surface area of the char. The rapid heating of the coal produces chars that have surface areas in the 200 m^2/g range which is orders of magnitude higher than the parent coal.

6.2.1.2 Form coke

Form coke would be used in the reduction of iron ore in blast furnaces. Coke is no longer used in the making of steel in Nova Scotia.

More than 50% of the coke ovens in the US are more than 20 years old and it is getting ever more difficult to meet environmental regulations. Despite the turndown in the production of pig iron in the US, the market for imported coke will increase due to the disproportionate decrease in the coke production. Imports in the 10 million tonnes/year range are needed.

Wooten et al., 1988 has reviewed six form coke production processes:

1. Eschweiller Bergwerkverein (Ancit), Germany
2. Bergbau-Forschung-Lurgi (BFL), Germany
3. Houilleres du Bassin de Nord (HBN), France
4. Sapozhinkov-Ukranian Institute of Coal Research, USSR
5. Didier-Keihan Rentan-Sumitomo Metal (DKS), Japan
6. FMC Corporation (FMC), United States

All of the processes carry out the entire form coke production.

The Ancit (Dartnell 1978, Hermann et al., 1977) process was operated at 275 t/d. The form coke has abrasion and strength problems which produced discouraging results.

The BFL (Dartnell 1978, Peters 1976) process was scaled up to 300 t/d but had operating and mechanical problems and was shut down.

The HBN (Cordier et al., 1977) process was operated at 100 t/d and tested in a blast furnace where 25% of the coke was form coke. A 200,000 t/yr plant has been designed but not built.

The Sapozhinkov (Dartnell 1978, Baker 1967) process was operated at 192 t/d. The form coke was used in small blast furnaces with good operating results.

The DKS (Wagener et al., 1977) process from coke has been used in a 50% blend with good results. A 10 oven experimental battery has been operated.

The FMC (Work 1966) process form coke has been tested in numerous blast furnace operations.

The processes listed above use different combinations of coking and non-coking coals and produce the char under various operating conditions. Testing of the form coke produced from the IGT process must be carried out in order to determine the suitability of the product in the metallurgical process.

Metallurgical coke typically has sulphur at < 1%, ash < 9% and volatile matter < 1%. The calculated char characteristics shown in Table 6.7 indicate that the char would have a sulphur at approximately 4%. This could be reduced by varying the feedstock and blending with a lower sulphur coking coal in the form coke step. Sulphur sorbents could also be added during the forming process. The ash in the char at 10% would need only a slight reduction. The char volatile matter is projected to be in the 9 - 16% range. This would be reduced during the briquetting step in the form coke process.

The Westray coal would meet the sulphur concentration requirements but would require a reduction in the ash levels.

6.2.1.3 Smokeless Domestic Fuel

Forming and pretreatment as carried out for the form coke use would be needed to produce smokeless fuel. Some additives may be needed to react with the sulphur and to ensure that the fuel maintained its form during combustion.

The target market would be the areas using anthracite, char coal and firewood. Markets may exist in the US if legislation is introduced to restrict the emissions from domestic sources. Significant markets exist in Korea and other Asian countries.

6.2.2 Condensable Hydrocarbons Utilization Options

The condensable hydrocarbons can be divided into the following categories as described in Table 6.8.

Table 6.8

The Condensable Hydrocarbons From the IGT Gasification Process and Their Descriptions

Product	Description
Light Oils	The light oils cut will be distilled from the tar and scrubbed from the gas stream and have a boiling point < 182°C.
Middle Oils	The middle oils tar fraction will be distilled in a temperature range between 182 and 310°C.
Heavy Oils	The heavy oils tar fraction will have a distillation range between 310 and 400°C.
Medium-Soft Pitch	The medium-soft pitch wili be the untreated distillation residue from the tar fractionation.
Medium-Hard Pitch	The medium-hard pitch will be produced as a result of the thermal treatment of the medium-soft pitch at 800°C.
Pitch Coke	The solid carbonized product produced from the thermal treatment of the medium-soft pitch will be the pitch coke.

The yields of oils and pitch from the IGT PRU tests are shown in Figure 6.4.

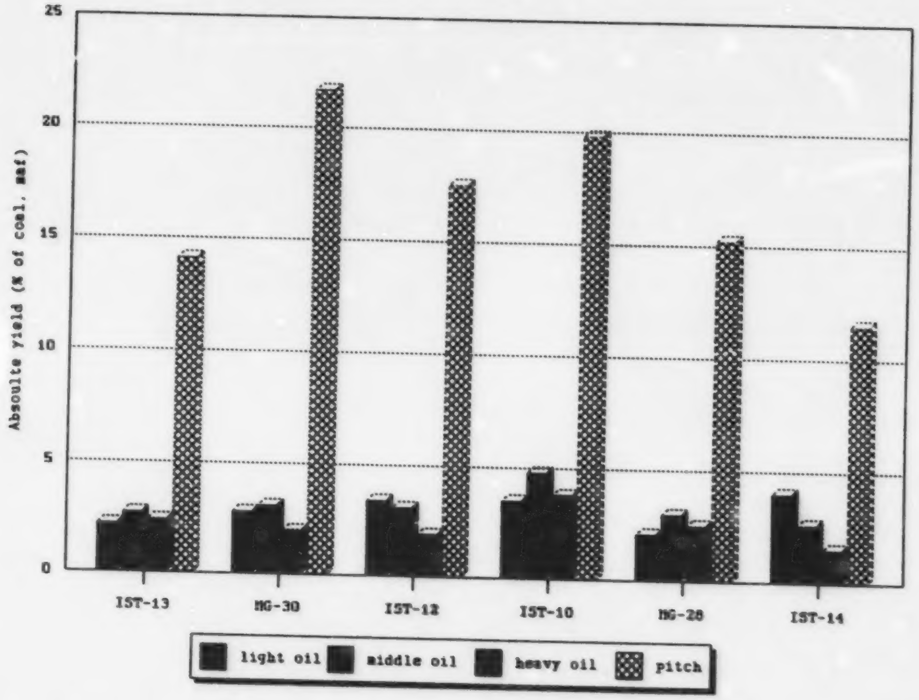


Figure 6.4

The Yields of Oils and Pitch From the IGT PRU Tests on West Virginia Coal

Condensable Hydrocarbon Uses

The light and medium oil fractions can be marketed to the chemical refining industry to make benzene, toluene, xylenes (BTX), phenol, cresols, xylenols, cresylic acids, naphthalene, indene and pyridine bases.

Severe refining conditions will be needed to remove sulphur, nitrogen and oxygen. The light oil fractions could produce a BTX mixture suitable as a gasoline octane enhancer. Limitations on the aromatic content in domestic fuels may inhibit the application of these chemicals as octane enhancers.

Phenols are used in the production of fireproof phenolic foam insulation. Naphthalene is converted to phthalic anhydride which is used as a plasticizer.

The heavy and middle oil fractions consist of creosote oil that could be used as a blending feedstock for timber preservative. Environmental concerns may limit the use of this product. The heavier ends of the creosote oil fractions can be used to produce carbon black.

The medium-soft pitch can be used as binders in road and roof applications.

Thermal upgrading of the medium-soft pitch would produce a product that would be used to produce carbon electrodes which are used in the aluminum and steel industries.

Tables 3.48 and 3.58 show the elemental and simulated distillation data for the oils/tars product (from the slipstream samples) for the IGT PRU tests using the WV coal.

The data in Tables 3.48 and 3.58 do not include the light oil fraction that boils below 149°C.

In all cases the oxygen content is nearly 10%. This property will have two results of significance: (1) it typically results in re-polymerization prior to hydrotreating if the oil is stored for any length of time and (2) the removal of the oxygen is the most difficult of any of the heteroatoms which requires the use of large quantities of hydrogen in hydrotreating processes. The sulphur content ranges from 0.5 to nearly 0.9% and is a function of: the parent coal % sulphur, the sulphur forms in the parent coal, the fluidizing gas composition, the bed material composition and the gasification conditions. Hydrotreating of the oils to remove the sulphur would be necessary prior to utilization.

6.2.3 Gas Utilization

The majority of the gas produced will be used on site to provide energy for the gasification reactor through the generation of hot flue gases used for fluidization. Part of the gas may be recycled to the reactor.

Some gas may be burned to provide process steam that would be used for coal drying and preheating, fractional distillation of the condensables and pitch cracking.

Gas produced in the IGT process will be used to provide process heat. Figure 6.5 shows a graphical representation of the IGT process gas yield composition on an N_2 free basis. H_2S has not been included in the graphical presentation. All of the components shown in Figure 6.5 would be acceptable for utilization in a process heat application. The H_2S would contribute the SO_2 levels shown in Table 6.9 if not removed prior to the combustion of the gas.

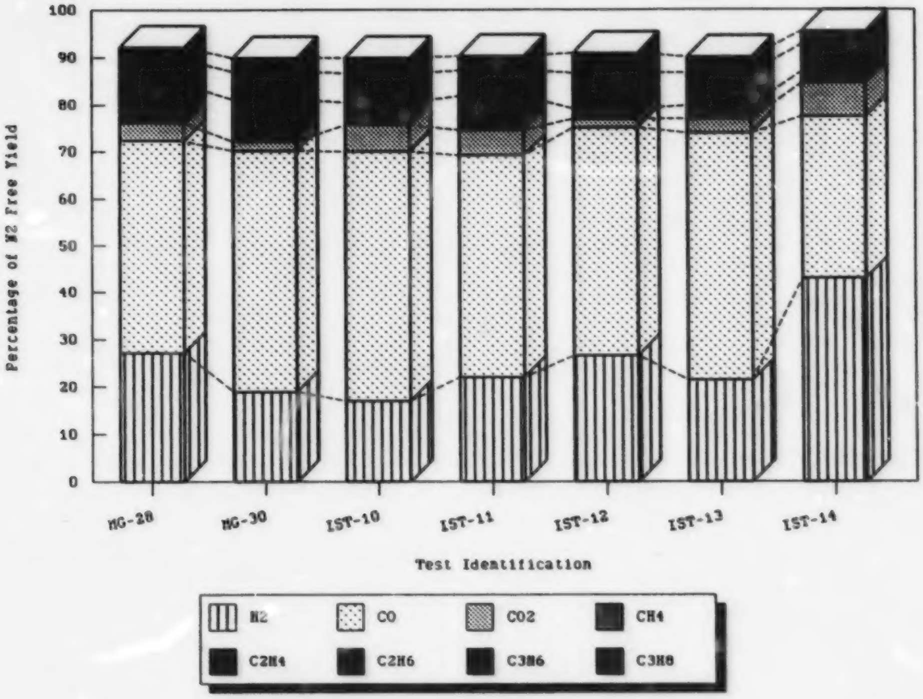


Figure 6.5

Gas Yield Composition From the IGT PRU Process
(N_2 Free Basis)

Table 6.9

SO_2 Released As a Result of Combustion of the Gas Generated From the IGT Process

Test ID	% H_2S	HHV (MJ/m^3)	g SO_2/MJ
MG-28	1.4	30.9	1.3
MG-30	1.6	34.3	1.3
IST-10	1.4	35.6	1.1
IST-11	0.9	33.3	0.8
IST-12	0.4	36.7	0.3
IST-13	1.4	36.2	1.1
IST-14	0.6	27.7	0.6

Variations in the release rate of SO_2 as presented in Table 6.9 are a function of the concentration of the H_2S and the heating values reported. Variations in the H_2S concentrations are somewhat as a result of the reaction of the gas with the reactor material. This would not be the case in a commercial operation. The use of recycled coal derived gas may result in variations in the H_2S concentrations, as more or less of the sulphur in the coal is released. The composition of the gas and its heating value may vary as a result of using a recycled coal-derived char as a heat source and process gas for fluidization. The concentrations reported are on a N_2 free basis. Levels of SO_2 in the 1 g/MJ range might be expected without removal of the H_2S. The analysis of the coal gasified is shown in Table 3.45. The sulphur content is 1.3% with approximately 1/3 pyritic and the remainder organic. The Nova Scotia coals (analysis shown in Tables 3.4 and 3.8) have 0.8 and 4% sulphur. The former having the same ratio of pyritic to organic as the West Virginia and the latter having the reverse ratio. The Westray coal might have SO_2 emission levels that are similar to the West Virginia. The Cape Breton coals would have higher SO_2 emission levels. In addition, the sulphur form variation seen in the Cape Breton coals would have an effect on the sulphur release mechanism and concentration. The US New Source Performance Standards are 0.52 g/MJ. H_2S stripping followed by a Claus removal process could be used to remove the H_2S.

6.3 Economic Analysis

The data of Wooten et al., 1988 has been drawn upon extensively in preparing this section of the report. No attempts have been made to generate original capital cost data. Wooten used cost estimates that are approximations extrapolated from data of plants of similar design. An attempt has been made to prepare an economic evaluation that achieves accuracies in the ± 40% range.

6.3.1 Capital Costs

The information generated by Wooten has been inflated at 5% per year from the 1987 US$ values to mid 1991 values. Conversion to CDN$ was made by multiplying by a factor of 1.17. Table 6.10 shows the plant sections that were costed. Table 6.11 shows the additional items that are included in the capital cost estimate. A contingency of 40% of the value of the plant cost has been added. The initial catalyst and chemical charges have been estimated. The total capital value shown for the three process options has been used in the subsequent analysis.

6.3.2 Operating and Maintenance Costs

Operating costs have been estimated assuming 328.5 standard working days per year. Table 6.12 shows the operating costs for the three IGT conceptual plant options. Included in the variable operating costs are the coal (dry basis), coal fines from the wash plant, electric power and limestone. The base case cost has been assumed to be $65/t (Crandlemire 1991). The price paid for coal fines from the wash plant has been assumed to be equal to that paid for the coal. The cost of electric power is 44.9 mil/kWh. This is the rate that NSPC charges to customers that use between 250 and 4000 kVA. Limestone in Cape Breton has been priced at $15/t (Crandlemire 1991). It has been assumed that the solid waste from the plant (spent sulphur sorbent) will be disposed of on site at cost of $4/t (Crandlemire 1991).

The fixed operating and maintenance costs have been calculated by adjusting the data provided by Wooten.

6.3.3 Credits

The projected product slate credit values have been determined by updating the unit prices provided by Wooten. Table 6.13 lists the product and the credit values that have been used.

The August, 1991 Chemical Marketing Reporter was used to generate the credit values for the following: pitch (roofing), pitch coke (liquid), light oil, medium oil, creosote oil, phenols and ammonium sulphate.

Table 6.10

IGT Mild Gasification Co-Product Conceptual Process Design, Process Equipment Costs

Plant Section	(Mid 1991 CDN $*1E6)		
	Option I	Option II	Option III
	Total Plant Cost	Total Plant Cost	Total Plant Cost
Coal Supply	$5.1	$5.7	$4.0
Mild Gasification	$59.7	$59.7	$57.7
Liquids and Gas Recovery	$59.7	$56.9	$59.7
Char Handling	$6.8	$5.7	$4.0
Char Briquetting	-	$19.9	$19.9
Form Coke Calcining	-	$71.1	-
Form Coke or Briquette Handling	-	$5.7	$6.0
Utilities and Offsites	$39.3	$59.7	$47.8
Total Plant Investment	$170.7	$284.4	$199.1

Table 6.11

IGT Mild Gasification Co-Product Conceptual Process Design, Capital Cost Summary

Summary of Capital Costs	(Mid 1991 CDN $*1E6)		
	Option I	Option II	Option III
Capital Equipment Costs	$170.7	$284.4	$119.1
Contingencies	$68.3	$113.8	$79.6
Initial Catalyst and Chemical Charges	$1.0	$1.0	$1.0
Prepaid Royalties	$1.4	$1.4	$1.4
Reproduction (Start-up) Costs	$14.2	$14.2	$14.2
Other Owner's Costs	$12.8	$12.8	$12.8
Working Capital	$12.8	$12.8	$12.8
Land	$1.4	$1.4	$1.4
Total Capital	$282.6	$441.9	$322.4

Table 6.12

IGT Mild Gasification Co-Product Conceptual Process Design, Operating and Maintenance Costs

OPERATING AND MAINTENANCE COSTS							
		Option I		Option II		Option III	
Item	Value	t/std. day	$*1E6/yr	t/std. day	$*1E6/yr	t/std. day	$*1E6/yr
VARIABLE OPERATING COSTS/EXPENSES							
Purchased Commodities							
Coal (dry basis)	65 $/t	7257	$155.0	2722	$58.1	2722	$58.1
Coal Fines	65 $/t	0	$0.0	1170	$25.0	1176	$25.1
Electric Power	44.9 mil/kwh$	3630 kw	$1.3	10670 kw	$3.8	8900 kw	$3.2
Limestone	15 /t	568	$2.8	568	$2.8	568	$2.8
Waste Disposal	4	772	$1.0	772	$1.0	772	$1.0
Catalyst and Chemicals			$0.7		$0.7		$0.7
Miscellaneous Supplies			$0.3		$0.3		$0.3
TOTAL VARIABLE COSTS			$161.0		$92.0		$91.0
FIXED OPERATING AND MAINTENANCE							
Labour							
Plant Management & Admin. Serv.			$1.7		$1.7		$1.7
Technical Service			$0.6		$0.6		$0.6
Plant Operations			$4.1		$4.6		$4.3
Plant Maintenance			$3.8		$4.3		$4.1
Maintenance Material			$2.0		$2.7		$2.4
General Corporate Expenses			$0.7		$0.7		$0.7
TOTAL FIXED COSTS			$13.0		$15.0		$14.0
PROPERTY TAXES AND INSURANCE			$2.4		$4.1		$3.0
TOTAL OPERATING COSTS			$176.4		$110.3		$108.0

Table 6.13

IGT Mild Gasification Co-Product Conceptual Process Design, Product Credit

Item	Unit Prices (/kg)	Value ($/t)	Option I		Option II		Option III	
			t/std. day	$*1E6/yr	t/std. day	$*1E6/yr	t/std. day	$*1E6/yr
PRODUCT CREDITS								
Pitch (roofing)	$0.361	$360.88	49	$5.8	73	$8.7	49	$5.8
Pitch Coke (liquid)	$0.379	$379.45	33	$4.1	49	$6.1	24	$2.9
Light Oil	$0.284	$283.54	69	$6.4	103	$9.6	70	$6.5
Medium Oil	$0.059	$59.39	200	$3.9	205	$4.0	200	$3.9
Creosote Oil	$0.235	$234.68	260	$20.1	284	$21.9	260	$20.1
Phenols	$0.709	$709.34	1	$0.3	1	$0.3	1	$0.3
Ammonium Sulphate	$0.072	$71.58	7	$0.2	7	$0.2	7	$0.2
Char/Coal Blend		$65.00	6180	$132.0	-	-	-	-
Form Coke		$122.85	-	-	2295	$92.6	-	-
Domestic Fuel		$91.78	-	-	-		2706	$81.6
TOTAL CREDITS				$172.6		$143.4		$121.2

The light oil fraction (benzene, toluene and xylene) would require some hydrotreating to remove sulphur. This would reduce the value of the product that would be produced from the gasification process. The value applied to this product slate will be 80% of the average of the Chemical Marketing Reporter values.

The price of naphthalene was used as representative of the medium oil product.

The value of the char/coal blend is assumed to equal that paid for the coal ($65/t).

The world price of metallurgical coke has been varying considerably over the past two years due to the change in the market demand. A value of $122.85/t has been used (Crandlemire 1991).

For the case of this evaluation, a price for domestic fuel has been determined based on the current approximate price of firewood ($91.78/t, air dry hardwood equivalent).

6.3.4 Cash Flow Analysis

The annual cash flow projections (shown in Table 6.14, 6.15 and 6.16) have been made for the three plant options. The book life of the plant has been taken equal to 30 years. The return on debt and equity (ROI) financing has been assumed to be equal. As other costs have been manipulated, the ROI has been varied in an iterative calculation procedure so that the net present value of the sum of the net profit (after tax) is equal to zero. The net profit figure includes payments of all taxes.

A gross profit tax rate of 50% has been used. It has been assumed that losses in the gross profit figures may be cumulatively carried forward and subtracted from the gross profit gains. Income taxes are not paid until the cumulative sum of the gross profit is greater than zero.

The annual inflation rate used is 5% (including escalation).

The financing of the capital cost has been divided into three categories: equity at 45%, debt at 35%, government (grant or investment at 0% ROI) at 20%.

A salvage value of 10% of the original capital cost (inflated to a future value) has been added to income in the final year of operation.

The debt is repaid on a straight line declining balance basis over the book life. The return on the debt is at fixed rate and in proportion to the outstanding debt amount.

The equity is repaid as a lump sum in the final year of operation.

The plant is assumed to be at 30% of full scale production in year three after the start of construction and 60% in year four and 100% in year five. The inflated credit values are weighted according to the percent of full scale operation.

Table 6.14

IGT Mild Gasification, Cash Flow Over Plant Life for Char/Coal Option

(all $ values in millions)

Year	Outstanding Capital Debt	% Full Operation	Credit Value	Expenses		Carrying Charges (CC)					Net Profit
				Fixed Operation & Maintenance Cost	Variable Operation & Maintenance Cost	Book Depreciation St. Line	Return on Debt	Return on Equity	Property Tax & Insurance	Income Tax	
1991	$49.5	0%	$0.0	$13.6	$0.0	$0.0	$0.0	$0.0	$2.5	$0.0	-
1992	$79.1	0%	$0.0	$14.3	$0.0	$0.0	$0.0	$0.0	$2.7	$0.0	-
1993	$98.9	30%	$60.0	$15.0	$55.9	$0.0	$0.0	$0.0	$2.8	$0.0	-
1994	$95.6	60%	$125.9	$15.7	$117.5	$3.3	$0.0	$0.0	$2.9	$0.0	-
1995	$92.3	100%	$220.3	$16.5	$205.6	$3.3	$0.0	$0.0	$3.1	$0.0	-
1996	$89.0	100%	$231.4	$17.3	$215.8	$3.3	$0.0	$0.0	$3.2	$0.0	-
1997	$85.7	100%	$242.9	$18.2	$226.6	$3.3	$0.0	$0.0	$3.4	$0.0	-
1998	$82.4	100%	$255.1	$19.1	$238.0	$3.3	$0.0	$0.0	$3.6	$0.0	-
1999	$79.1	100%	$267.8	$20.1	$249.9	$3.3	$0.0	$0.0	$3.8	$0.0	-
2000	$75.8	100%	$281.2	$21.1	$262.3	$3.3	$0.0	$0.0	$3.9	$0.0	-
2001	$72.5	100%	$295.3	$22.1	$275.5	$3.3	$0.0	$0.0	$4.1	$0.0	-
2002	$69.2	100%	$310.0	$23.2	$289.2	$3.3	$0.0	$0.0	$4.3	$0.0	-
2003	$65.9	100%	$325.6	$24.4	$303.7	$3.3	$0.0	$0.0	$4.6	$0.0	-
2004	$62.6	100%	$341.8	$25.6	$318.9	$3.3	$0.0	$0.0	$4.8	$0.0	-
2005	$59.3	100%	$358.9	$26.9	$334.8	$3.3	$0.0	$0.0	$5.0	$0.0	-
2006	$56.0	100%	$376.9	$28.2	$351.6	$3.3	$0.0	$0.0	$5.3	$0.0	-
2007	$52.7	100%	$395.7	$29.7	$369.1	$3.3	$0.0	$0.0	$5.5	$0.0	-
2008	$49.5	100%	$415.5	$31.1	$387.6	$3.3	$0.0	$0.0	$5.8	$0.0	-
2009	$46.2	100%	$436.3	$32.7	$407.0	$3.3	$0.0	$0.0	$6.1	$0.0	-
2010	$42.9	100%	$458.1	$34.3	$427.3	$3.3	$0.0	$0.0	$6.4	$0.0	-
2011	$39.6	100%	$481.0	$36.1	$448.7	$3.3	$0.0	$0.0	$6.7	$0.0	-

Table 6.14 (Continued)

IGT Mild Gasification, Cash Flow Over Plant Life for Char/Coal Option

(all $ values in millions)

Year	Outstanding Capital Debt	% Full Operation	Credit Value	Expenses		Carrying Charges (CC)					Net Profit
				Fixed Operation & Maintenance Cost	Variable Operation & Maintenance Cost	Book Depreciation St. Line	Return on Debt	Return on Equity	Property Tax & Insurance	Income Tax	
2012	$36.3	100%	$505.0	$37.9	$471.1	$3.3	$0.0	$0.0	$7.1	$0.0	-
2013	$33.0	100%	$530.3	$39.8	$494.7	$3.3	$0.0	$0.0	$7.4	$0.0	-
2014	$29.7	100%	$556.8	$41.7	$519.4	$3.3	$0.0	$0.0	$7.8	$0.0	-
2015	$26.4	100%	$584.6	$43.8	$545.4	$3.3	$0.0	$0.0	$8.2	$0.0	-
2016	$23.1	100%	$613.9	$46.0	$572.7	$3.3	$0.0	$0.0	$8.6	$0.0	-
2017	$19.8	100%	$644.6	$48.3	$601.3	$3.3	$0.0	$0.0	$9.0	$0.0	-
2018	$16.5	100%	$676.8	$50.7	$631.4	$3.3	$0.0	$0.0	$9.5	$0.0	-
2019	$13.2	100%	$710.6	$53.3	$662.9	$3.3	$0.0	$0.0	$10.0	$0.0	-
2020	$9.9	100%	$746.2	$55.9	$696.1	$3.3	$0.0	$0.0	$10.4	$0.0	-
2021	$6.6	100%	$783.5	$58.7	$730.9	$3.3	$0.0	$0.0	$11.0	$0.0	-
2022	$3.3	100%	$822.7	$61.7	$767.4	$3.3	$0.0	$0.0	$11.5	$0.0	-
2023	$0.0	100%	$863.8	$64.7	$805.8	$3.3	$0.0	$127.2	$12.1	$0.0	-

Table 6.15

IGT Mild Gasification, Cash Flow Over Plant Life for Form Coke Option

(all $ values in millions)

Year	Outstanding Capital Debt	% Full Operation	Credit Value	Expenses		Carrying Charges (CC)					Net Profit
				Fixed Operation & Maintenance Cost	Variable Operation & Maintenance Cost	Book Depreciation St. Line	Return on Debt	Return on Equity	Property Tax & Insurance	Income Tax	
1991	$77.3	0%	$0.0	$15.2	$0.0	$0.0	$13.4	$26.8	$4.3	$0.0	($59.7)
1992	$123.7	0%	$0.0	$16.0	$0.0	$0.0	$21.4	$26.8	$4.5	$0.0	($68.7)
1993	$154.7	30%	$49.8	$16.8	$31.8	$0.0	$26.8	$26.8	$4.8	$0.0	($57.1)
1994	$149.5	60%	$104.6	$17.6	$66.9	$5.2	$25.9	$26.8	$5.0	$0.0	($42.7)
1995	$144.3	100%	$183.0	$18.5	$117.0	$5.2	$25.0	$26.8	$5.3	$0.0	($14.7)
1996	$139.2	100%	$192.2	$19.4	$122.9	$5.2	$24.1	$26.8	$5.5	$0.0	($11.7)
1997	$134.0	100%	$201.8	$20.4	$129.0	$5.2	$23.2	$26.8	$5.8	$0.0	($8.6)
1998	$128.9	100%	$211.9	$21.4	$135.5	$5.2	$22.3	$26.8	$6.1	$0.0	($5.4)
1999	$123.7	100%	$222.5	$22.5	$142.2	$5.2	$21.4	$26.8	$6.4	$0.0	($2.0)
2000	$118.6	100%	$233.6	$23.6	$149.3	$5.2	$20.5	$26.8	$6.7	$0.0	$1.4
2001	$113.4	100%	$245.3	$24.8	$156.8	$5.2	$19.6	$26.8	$7.1	$0.0	$5.0
2002	$108.3	100%	$257.5	$26.1	$164.7	$5.2	$18.7	$26.8	$7.4	$0.0	$8.7
2003	$103.1	100%	$270.4	$27.4	$172.9	$5.2	$17.8	$26.8	$7.8	$0.0	$12.6
2004	$97.9	100%	$283.9	$28.7	$181.5	$5.2	$17.0	$26.8	$8.2	$0.0	$16.6
2005	$92.8	100%	$298.1	$30.2	$190.6	$5.2	$16.1	$26.8	$8.6	$0.0	$20.8
2006	$87.6	100%	$313.0	$31.7	$200.1	$5.2	$15.2	$26.8	$9.0	$0.0	$25.1
2007	$82.5	100%	$328.7	$33.2	$210.1	$5.2	$14.3	$26.8	$9.5	$0.0	$29.6
2008	$77.3	100%	$345.1	$34.9	$220.6	$5.2	$13.4	$26.8	$9.9	$0.0	$34.3
2009	$72.2	100%	$362.3	$36.7	$231.7	$5.2	$12.5	$26.8	$10.4	$0.0	$39.2
2010	$67.0	100%	$380.5	$38.5	$243.3	$5.2	$11.6	$26.8	$10.9	$0.0	$44.2
2011	$61.9	100%	$399.5	$40.4	$255.4	$5.2	$10.7	$26.8	$11.5	$24.8	$24.8

Table 6.15 (Continued)

IGT Mild Gasification, Cash Flow Over Plant Life for Form Coke Option

(all $ values in millions)

Year	Outstanding Capital Debt	% Full Operation	Credit Value	Expenses		Carrying Charges (CC)					Net Profit
				Fixed Operation & Maintenance Cost	Variable Operation & Maintenance Cost	Book Depreciation St. Line	Return on Debt	Return on Equity	Property Tax & Insurance	Income Tax	
2012	$56.7	100%	$419.5	$42.4	$268.2	$5.2	$9.8	$26.8	$12.1	$27.5	$27.5
2013	$51.6	100%	$440.4	$44.6	$281.6	$5.2	$8.9	$26.8	$12.7	$30.4	$30.4
2014	$46.4	100%	$462.5	$46.8	$295.7	$5.2	$8.0	$26.8	$13.3	$33.4	$33.4
2015	$41.2	100%	$485.6	$49.1	$310.5	$5.2	$7.1	$26.8	$14.0	$36.5	$36.5
2016	$36.1	100%	$509.9	$51.6	$326.0	$5.2	$6.2	$26.8	$14.7	$39.7	$39.7
2017	$30.9	100%	$535.4	$54.2	$342.3	$5.2	$5.4	$26.8	$15.4	$43.1	$43.1
2018	$25.8	100%	$562.1	$56.9	$359.4	$5.2	$4.5	$26.8	$16.2	$46.6	$46.6
2019	$20.6	100%	$590.2	$59.7	$377.4	$5.2	$3.6	$26.8	$17.0	$50.3	$50.3
2020	$15.5	100%	$619.7	$62.7	$396.3	$5.2	$2.7	$26.8	$17.8	$54.2	$54.2
2021	$10.3	100%	$650.7	$65.8	$416.1	$5.2	$1.8	$26.8	$18.7	$58.2	$58.2
2022	$5.2	100%	$683.3	$69.1	$436.9	$5.2	$0.9	$26.8	$19.7	$62.4	$62.4
2023	$0.0	100%	$717.4	$72.6	$458.7	$5.2	$0.0	$181.4	$20.6	$0.0	($21.1)

Table 6.16

IGT Mild Gasification, Cash Flow Over Plant Life for Smokeless Fuel Option

(all $ values in millions)

Year	Outstanding Capital Debt	% Full Operation	Credit Value	Expenses		Carrying Charges (CC)					Net Profit
				Fixed Operation & Maintenance Cost	Variable Operation & Maintenance Cost	Book Depreciation St. Line	Return on Debt	Return on Equity	Property Tax & Insurance	Income Tax	
1991	$56.4	0%	$0.0	$14.5	$0.0	$0.0	$4.1	$8.1	$3.1	$0.0	($29.8)
1992	$90.3	0%	$0.0	$15.2	$0.0	$0.0	$6.5	$8.1	$3.3	$0.0	($33.2)
1993	$112.8	30%	$42.1	$16.0	$31.7	$0.0	$8.1	$8.1	$3.5	$0.0	($25.3)
1994	$109.1	60%	$88.4	$16.8	$66.5	$3.8	$7.9	$8.1	$3.6	$0.0	($18.3)
1995	$105.3	100%	$154.7	$17.6	$116.4	$3.8	$7.6	$8.1	$3.8	$0.0	($2.6)
1996	$101.6	100%	$162.5	$18.5	$122.2	$3.8	$7.3	$8.1	$4.0	$0.0	($1.4)
1997	$97.8	100%	$170.6	$19.4	$128.3	$3.8	$7.1	$8.1	$4.2	$0.0	($0.3)
1998	$94.0	100%	$179.1	$20.4	$134.7	$3.8	$6.8	$8.1	$4.4	$0.0	$0.9
1999	$90.3	100%	$188.1	$21.4	$141.4	$3.8	$6.5	$8.1	$4.6	$0.0	$2.2
2000	$86.5	100%	$197.5	$22.5	$148.5	$3.8	$6.2	$8.1	$4.9	$0.0	$3.5
2001	$82.8	100%	$207.4	$23.6	$155.9	$3.8	$6.0	$8.1	$5.1	$0.0	$4.8
2002	$79.0	100%	$217.7	$24.8	$163.7	$3.8	$5.7	$8.1	$5.4	$0.0	$6.2
2003	$75.2	100%	$228.6	$26.0	$171.9	$3.8	$5.4	$8.1	$5.6	$0.0	$7.7
2004	$71.5	100%	$240.0	$27.3	$180.5	$3.8	$5.2	$8.1	$5.9	$0.0	$9.2
2005	$67.7	100%	$252.0	$28.7	$189.5	$3.8	$4.9	$8.1	$6.2	$0.0	$10.8
2006	$63.9	100%	$264.6	$30.1	$199.0	$3.8	$4.6	$8.1	$6.5	$0.0	$12.5
2007	$60.2	100%	$277.9	$31.6	$209.0	$3.8	$4.3	$8.1	$6.8	$0.0	$14.2
2008	$56.4	100%	$291.8	$33.2	$219.4	$3.8	$4.1	$8.1	$7.2	$0.0	$16.0
2009	$52.7	100%	$306.4	$34.9	$230.4	$3.8	$3.8	$8.1	$7.5	$0.0	$17.8
2010	$48.9	100%	$321.7	$36.6	$241.9	$3.8	$3.5	$8.1	$7.9	$9.9	$9.9
2011	$45.1	100%	$337.8	$38.4	$254.0	$3.8	$3.3	$8.1	$8.3	$10.9	$10.9

Table 6.16 (Continued)

IGT Mild Gasification, Cash Flow Over Plant Life for Smokeless Fuel Option

(all $ values in millions)

Year	Outstanding Capital Debt	% Full Operation	Credit Value	Expenses		Carrying Charges (CC)					Net Profit
				Fixed Operation & Maintenance Cost	Variable Operation & Maintenance Cost	Book Depreciation St. Line	Return on Debt	Return on Equity	Property Tax & Insurance	Income Tax	
2012	$41.4	100%	$354.6	$40.4	$266.7	$3.8	$3.0	$8.1	$8.7	$12.0	$12.0
2013	$37.6	100%	$372.4	$42.4	$280.0	$3.8	$2.7	$8.1	$9.2	$13.1	$13.1
2014	$33.9	100%	$391.0	$44.5	$294.0	$3.8	$2.4	$8.1	$9.6	$14.2	$14.2
2015	$30.1	100%	$410.5	$46.7	$308.7	$3.8	$2.2	$8.1	$10.1	$15.4	$15.4
2016	$26.3	100%	$431.1	$49.0	$324.2	$3.8	$1.9	$8.1	$10.6	$16.7	$16.7
2017	$22.6	100%	$452.6	$51.5	$340.4	$3.8	$1.6	$8.1	$11.1	$18.0	$18.0
2018	$18.8	100%	$475.3	$54.1	$357.4	$3.8	$1.4	$8.1	$11.7	$19.4	$19.4
2019	$15.0	100%	$499.0	$56.8	$375.3	$3.8	$1.1	$8.1	$12.3	$20.8	$20.8
2020	$11.3	100%	$524.0	$59.6	$394.0	$3.8	$0.8	$8.1	$12.9	$22.3	$22.3
2021	$7.5	100%	$550.2	$62.6	$413.7	$3.8	$0.5	$8.1	$13.6	$23.9	$23.9
2022	$3.8	100%	$577.7	$65.7	$434.4	$3.8	$0.3	$8.1	$14.2	$25.5	$25.5
2023	$0.0	100%	$606.6	$69.0	$456.2	$3.8	$0.0	$121.0	$14.9	$0.0	($58.3)

6.3.5 Cost Sensitivity Analysis

A number of cost parameters were varied to evaluate their economic sensitivity. The "base case" economic data are presented in Tables 6.10, 6.11, 6.12 and 6.13. The effect of varying individual costs as the others are held at the "base case" are discussed below.

ROI Sensitivity to Capital Cost

The capital cost section of the analysis has the least accuracy. The base case data contains a contingency to compensate for the lack of accuracy valued at 40% of the plant capital costs. The results of the sensitivity of the ROI to the variations in the overall capital costs are shown in Figure 6.6. In all cases the ROI does not show a dramatic response to variations in the capital cost. This lack of sensitivity reduces the effect of the projected ± 40% accuracy of the study.

The char/coal blend option cannot show a positive ROI under any of the capital costs considered.

The form coke scenario projects the most favourable ROI as the capital costs are varied from the base case. ROI's vary from 24 to 13% as the capital cost is varied by -25% and +25% respectively.

The smokeless fuel option shows a marginal ROI (7%) under the base case situation. The ROI changes from 10% to 5% as the capital cost is varied by 25% (down and up respectively).

ROI Sensitivity to Coal Cost

Production costs result in Nova Scotia coal being relatively expensive (base case price of $65/t). Coal costs have been varied down from the base case in consideration that our coal is one of the most expensive (Figure 6.7).

The char/coal option shows little sensitivity to the coal cost and does not show a favourable ROI at any coal costs used.

Both the form coke and smokeless fuel options show significant sensitivity to variations in the coal price.

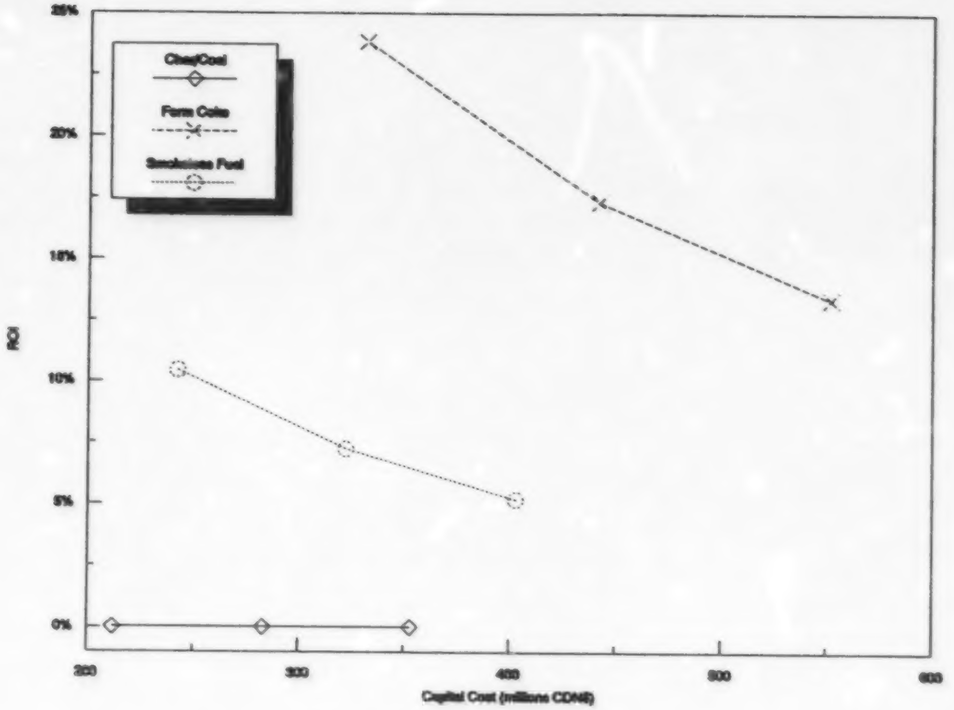


Figure 6.6

IGT Conceptual Process, Return on Investment Versus the Capital Cost

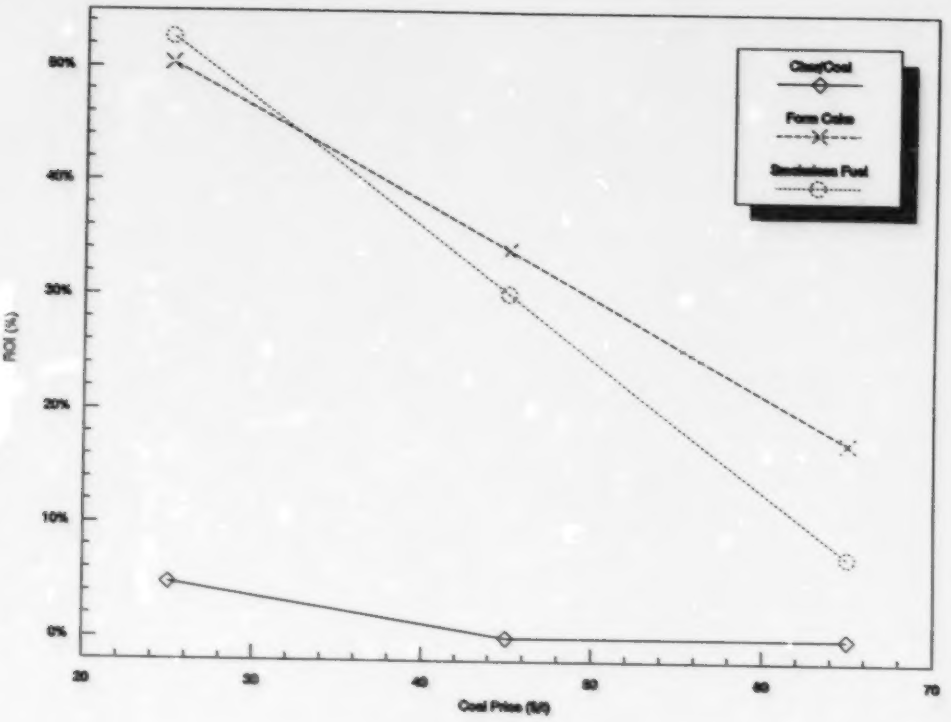


Figure 6.7

IGT Conceptual Process, Return on Investment Versus the Price Paid for Coal

Sensitivity to Product Value

Figure 6.8 shows the relative value of the products for the three process options. The relative value will remain constant if the value of the products inflate at the same rate over the plant life. The sensitivity of the ROI to the char based product has been evaluated due to its dominant relative value in all three options.

Figure 6.9 shows the ROI sensitivity to the value of the major product, char, form coke or smokeless fuel. The sensitivity of the ROI to product value decreases from the char/coal option to the smokeless fuel option and to the form coke option. Recent variations in the value of metallurgical coke increases the significance of the lower sensitivity to the product value.

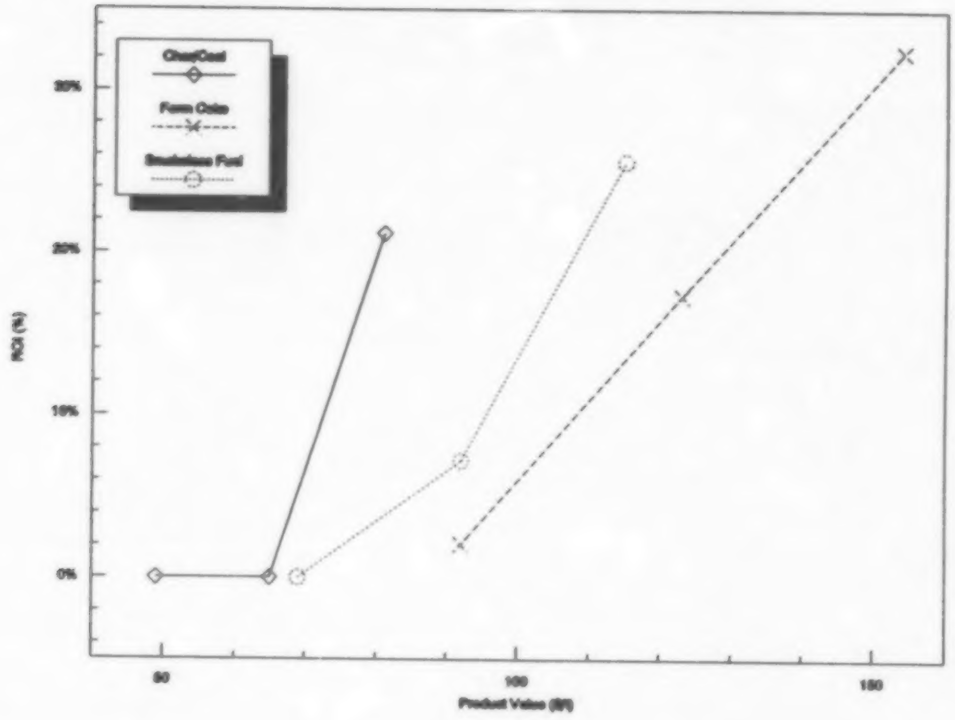


Figure 6.8

IGT Conceptual Process, Return on Investment Versus the Price of the Solid Product Value

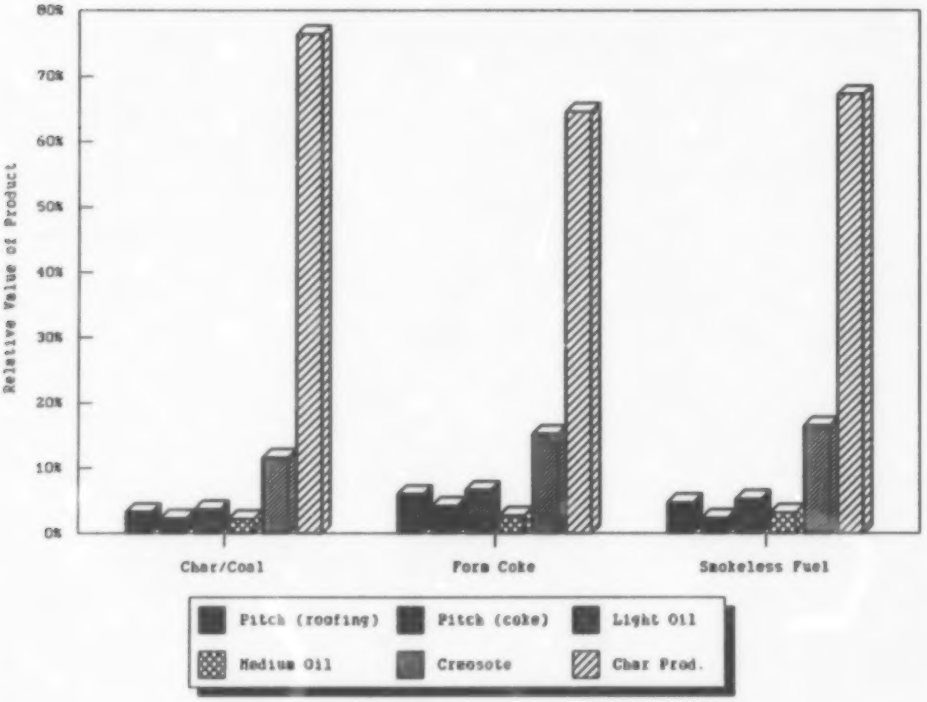


Figure 6.9

IGT Conceptual Process, Relative Value of Products

7 RECOMMENDATIONS

The technology development carried out on the VEW system should be monitored.

The potential of integrating a VEW gasifier with the 165 MWe CFBC at Point Aconi should be determined.

The promising technical and economic data on the Institute of Gas Technology co-product process suggests that further evaluation should be carried out.

The author recommends that further limited evaluation of the IGT process be done in conjunction with the IGT PDU operation. The performance of the soon to be operated IGT 1 t/h Process Development Unit (PDU) should be analyzed. The PDU will have, among other new steps, integrated hot char recirculation and a high temperature pitch hydrotreating step. The effect of these process modifications on the product yields and properties should be monitored. The economic evaluation should be updated as new data becomes available. The viability and costs of processing a Nova Scotia coal in the IGT PDU should be established.

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